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ACID MINE DRAINAGE TREATMENT BY ION EXCHANGE

J. G. Holmes, et al

Culligan International Company

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Acid Mine Drainage Treatment by Ion Exchange

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16. Abstract

Laboratory studies were conducted on synthetic acid mine drainage (AMD) using five ion exchange processes as follows: strong acid cation exchanger, hydrogen form; strong acid cation exchanger, sodium form; weak base anion exchanger, free base form; weak base anion exchanger, bicarbonate form; strong base anion exchanger, sulfate form.

Studies in the first stage eliminated two resins from further work: strong acid cation exchanger, sodium form; and strong base anion exchanger, sulfate form. The remaining three processes were studied additionally to establish fundamental design parameters for plants which can produce potable water from AMD.

Two processes have resulted. These were used as the basis for design of plants in three sizes (0.1, 0.5, and 1.0 MGD) of each process.

Cost estimates have been developed for operating costs, equipment (unassembled and unerected), erection costs based on electrical and plumbing requirements.

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ACID MINE DRAINAGE TREATMENT BY ION EXCHANGE

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Contract No. 14-12-887
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ABSTRACT

Laboratory studies were conducted on synthetic acid mine drainage treatment using ion exchange processes. These studies were in two stages. During the first stage, five representative ion exchange resins of the various types which are commercially available were surveyed through laboratory column test studies, to determine their applicability in the treatment of acid mine drainage (AMD). The second stage of the laboratory studies selected the three resins which were feasible in the treatment of AMD in the first stages. These three selected resins were then studied further in the treatment of synthetic AMD containing 100% ferrous iron, and in separate tests, 100% ferric iron and to study a total process for the production of potable water.

The resins studied in the first stage were as follows:

Strong acid cation exchanger regenerated with sulfuric acid.
Strong acid cation exchanger regenerated with sodium chloride.
Weak base anion exchanger regenerated with caustic soda.
Weak base anion exchanger regenerated with sodium hydroxide
  and carbon dioxide (modified Desal process).
Strong base anion exchanger operated as in the SUL-biSUL process.

The following two resins were eliminated from further consideration. The strong acid cation exchanger regenerated with sodium chloride produced an effluent which was increased in total dissolved solids: regeneration with sulfuric acid was a better process. The strong base anion exchanger operated as in the SUL-biSUL process produced the lowest volume of treated effluent per unit volume of exchanger.

The three recommended ion exchange resins were studied to establish fundimental design parameters for treatment plants. Process optimization was not attempted; rather, feasibility and basic parameters were established.

Based on the laboratory studies, two complete processes for the treatment of AMD by ion exchange techniques were established; the two resin system and the modified Desal system. Treatment plants in three sizes; 0.1, 0.5 and 1.0 MGD, were designed for each system so that cost estimates could be established. These estimates are presented in the report.

Continuation of the work is recommended in the form of pilot plant studies on the two resin system.

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Key words: Acid mine drainage, demineralization, ion exchange, cost.

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SECTION 1

CONCLUSIONS

Based on the experimental data obtained in this study, we conclude the following:

A potable water can be obtained from acid mine drainage by treatment methods which incorporate the use of specific ion exchange processes.

One specific process (2 resin system) which was found successful used a strong acid cation exchanger (hydrogen form), followed by a weak base anion exchanger (free base form) followed by post treatment consisting of pH elevation to 9.9, aeration, filtration and final pH correction. The chemical costs for producing water by this process were estimated at about 63¢/1000 gallons when using a particular synthetic AMD influent. Liquid wastes from this process would be acidic requiring subsequent treatment and disposal. Estimated unerected, uninstalled equipment costs for this process are $126,000 for a 0.1 MGD plant; $256,000 for a 0.5 MGD plant; $428,000 for a 1.0 MGD plant.

A second process (modified Desal(1) process) which was found successful used a weak base anion exchanger (Rohm and Haas IRA-68) in the bicarbonate form followed by aeration, lime treatment, filtration and final pH correction. The chemical costs for producing water by this process were estimated at about 48¢/1000 gallons when using a synthetic AMD influent similar to that used in the first treatment process described above. The wastes from this process would be an alkaline liquid and a lime treatment sludge. Estimated unerected, uninstalled equipment costs for this process are $156,000 for a 0.1 MGD plant; $323,000 for a 0.5 MGD plant; $465,000 for a 1.0 MGD plant.

The treatment process utilizing a strong acid cation exchanger in the sodium form was studied and found to be less efficient than the hydrogen form for the production of potable water from acid mine drainage.

The treatment process utilizing a strong base anion exchanger in the sulfate form (SUL-biSUL process(2)) was studied also. This process appeared feasible for the production of potable water if coupled to a strong acid cation exchanger (hydrogen form). However, because of the more promising outlook of the previously mentioned processes (above), this process was not selected for intensive study.

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SECTION 2

RECOMMENDATIONS

It is recommended that additional investigations be conducted to establish the best weak base anion exchanger for use in the strong acid cation (hydrogen form) - weak base anion (free base form) treatment process. This investigation should also include a study of resin stability and the rate of iron accumulation on various weak base anion exchangers when subjected to repeated cycles of regeneration and treatment with a simulated cation effluent. The new investigation must also study efficient methods for disposal of regenerant wastes. Analytical methods must be established to enable correct determination of free mineral acidity as it affects the ion exchange reactions.

After performing the study described above, a pilot plant should be constructed for the application of the complete process to an actual acid mine drainage source. The pilot plant should be designed and operated for the following purposes:

1. Establish the economic comparison of sulfuric acid and hydrochloric acid regenerations taking into consideration the volume of waste regenerant, its treatment and ultimate disposal.

2. Establish all operating costs for producing potable water by this complete treatment process. This will entail a consideration of expected ion exchanger life and the ability to restore the anion exchanger by iron removal processes.

3. Partially optimize the process to enable adequate design of larger plants.

4. Establish the economics of using hydrochloric acid and sulfuric acid (separately) as regenerants for cation exchange resins.

No recommendations are made relative to the weak base anion (bicarbonate form) - lime treatment process. This process will produce a potable effluent also. However, a plant utilizing this process is already under construction by the Commonwealth of Pennsylvania.

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SECTION 3

INTRODUCTION

The demands for potable water are steadily increasing. In areas where acid mine drainage has diminished the available supplies, processes which are capable of producing potable water from this drainage are of particular interest.

Ion exchange processes have this capability. Certain specialized ion exchange processes (modified Desal(1) and SUL-biSUL(2))have already been employed for the treatment of acid mine drainage. However, very little attention has been given to the application of conventional methods of ion exchange in solving this problem.

This investigation was intended to study several conventional ion exchange processes, using commercially available materials and to determine if any of these processes could be used to produce potable water from acid mine drainage.

All the laboratory studies were performed in the research laboratories of the Culligan International Company. Laboratory studies were started August 18, 1970 and were completed on August 27, 1971.

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SECTION 4

OBJECTIVES

This project was initiated to achieve the following objectives. The first objective was to study five different ion exchange processes for the treatment of acid mine drainage in order to determine which of the processes could be applied successfully to the treatment of acid mine drainage. One month was allotted for each process study.

A second objective was the further study of three processes selected from the original five. The three processes were to be selected on the basis of predictable success as an entire process or as a portion of an entire process for the production of potable water from AMD. Two months were allotted for each process study: one month was to be spent with ferrous iron, the second month was to be spent with ferric iron.

The third objective was to establish capital cost estimated for plants by the design of three complete treatment plants in two sizes each (1 MGD and 0.5 MGD) for the production of potable water from acid mine drainage using three different treatment processes. This was later revised to cover two complete treatment plants in three sizes each (1 MGD, 0.5 MGD and 0.1 MGD). This was done because two of the three processes selected for further study were really component steps of a single complete treatment process.

The accomplishment of these objectives provides a basis for determining the merit of installing large plants to treat AMD by ion exchange process. That is, the project compared the technical feasibility, operating costs and plant costs.

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SECTION 5

LABORATORY TEST APPARATUS DESCRIPTION

The ion exchange processes were studied in 2-inch ID by 60 inches acrylic columns. The acrylic columns were fitted at the lower end with a suitable strainer with openings small enough to prevent loss of the 10 by 20 mesh flint gravel support for the ion exchanger. A strainer was used at the top of the column when upflow treatment cycles were employed. This was done to prevent loss of ion exchange materials into the treated effluent.

Figures 1 and 2 are photographs showing the ion exchanger column apparatus with the two columns used. Figure 3 is a schematic diagram of the entire apparatus. The interconnecting piping consisted of 1/4-inch PVC pipe. The 500 gallon and 20 gallon reservoirs were fiberglass and polyethylene construction respectively. The pumps used to pressurize the solutions from the reservoirs were small, low head centrifugal pumps (Eastern Industries). Valving consisted of plastic ball valves of appropriate sizes arranged so fluids from either of the two reservoirs could be passed through the columns in an upflow or downflow direction. Fluids could alternately be passed through one column and then the next in either direction. An alternate connection was available for conducting any other source fluid (e.g., CO_2 saturated water) through the columns.

Flow indicator and control devices were plumbed into the piping at convenient locations so that fluid flow rates could be observed. The waste effluents were led to neutralizing equipment before being discharged to waste. Composite effluents were normally collected in polyethylene reservoirs before being mixed, sampled and discharged to waste.

The following reagent grade chemicals and demineralized water were used to prepare the synthetic acid mine drainage (AMD) solution.

TABLE 1. Reagent Chemicals Used to Prepare Synthetic AMD.

H_2SO_4
$CaSO_4 \cdot 2H_2O$
$MnSO_4 \cdot 1H_2O$
$Al_2(SO_4)_3 \cdot 18H_2O$
$MgSO_4 \cdot 7H_2O$
$FeSO_4 \cdot 7H_2O$
$Fe_2(SO_4)_3 \cdot XH_2O$

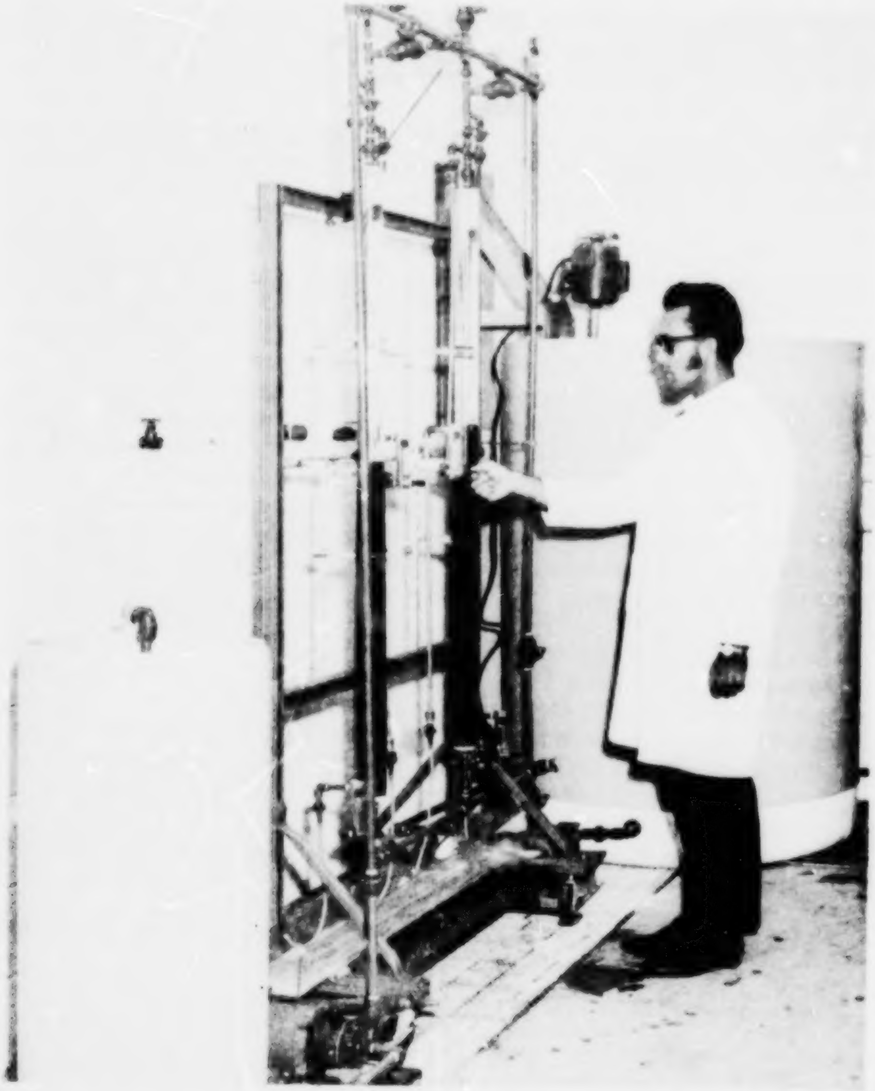
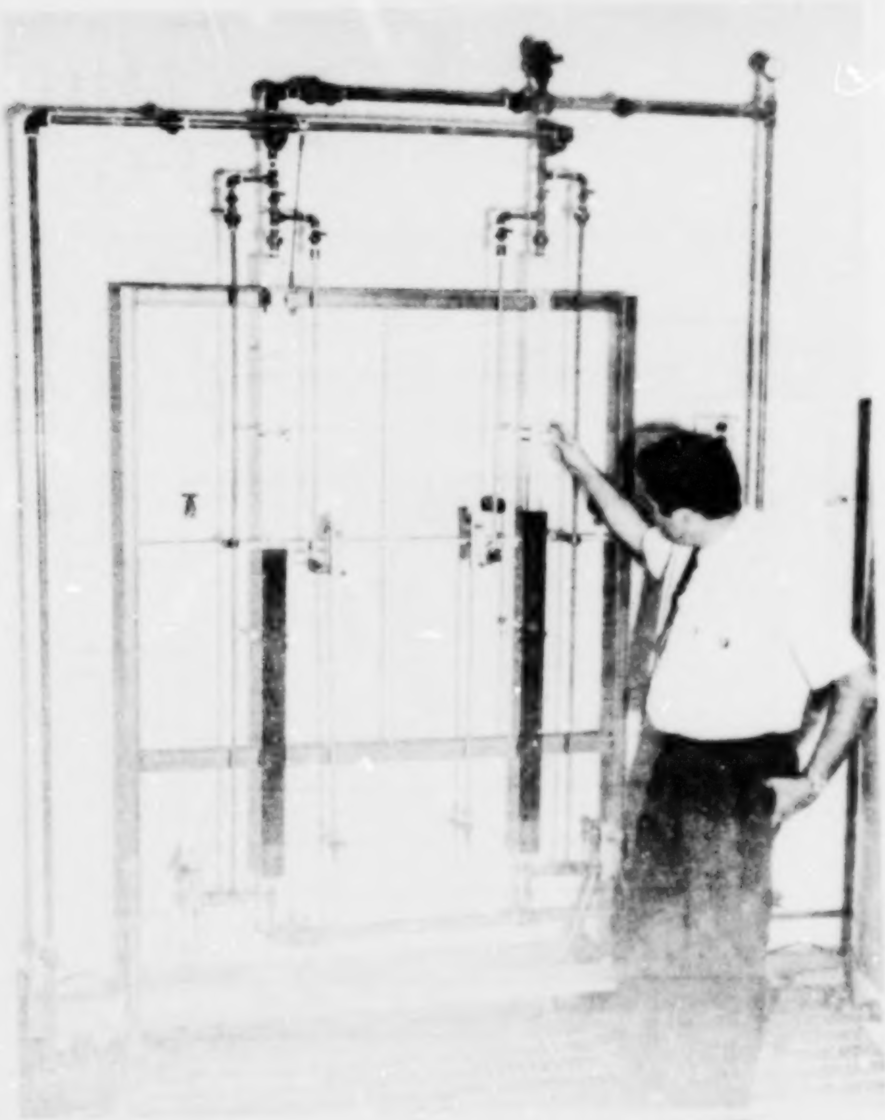


Figure 1. Acid Mine Drainage Treatment Test Apparatus



Reproduced from best available copy.

Figure 2. Ion Exchanger Columns

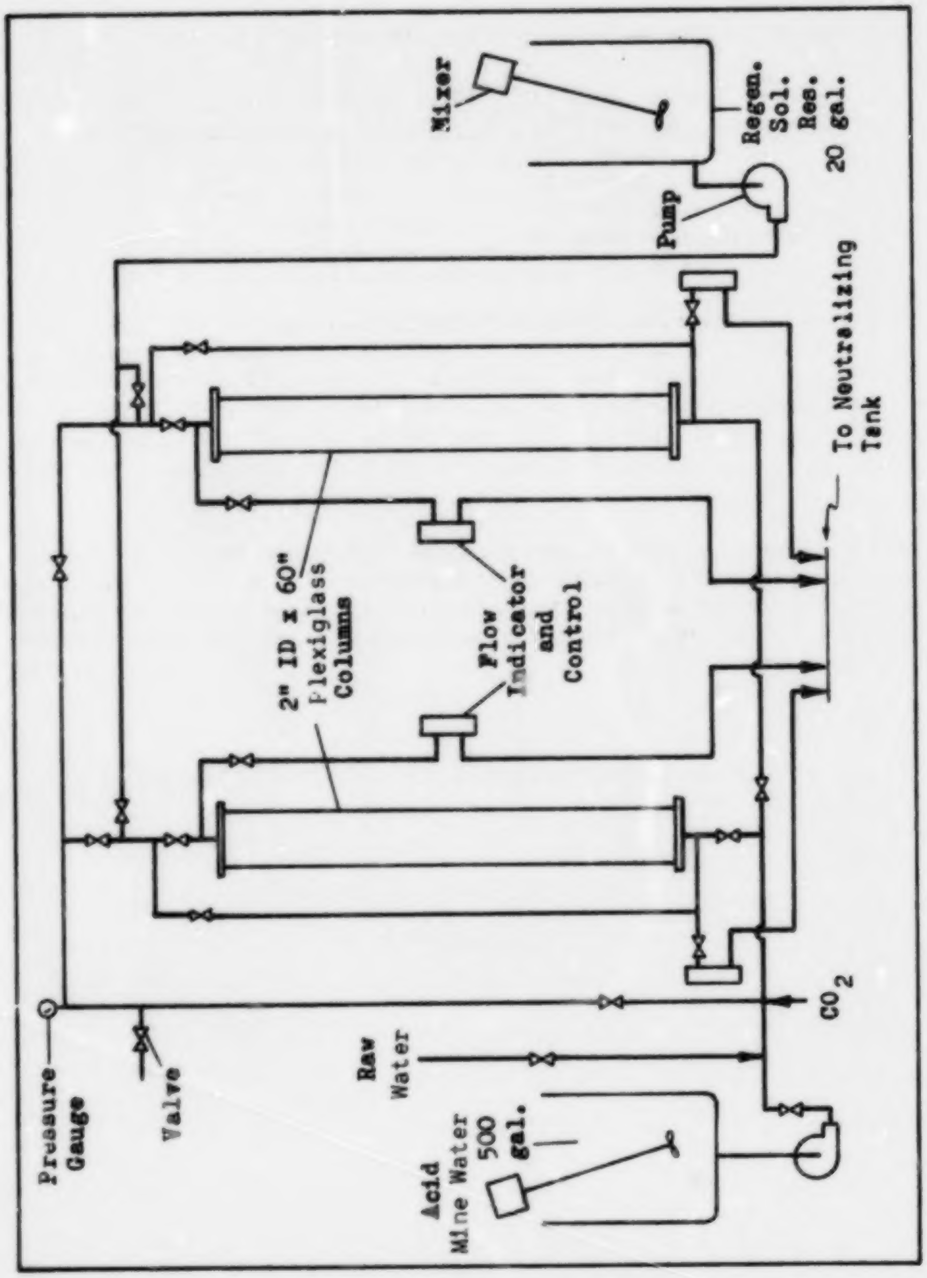

Pressure Gauge
Valve
Acid Mine Water 500 gal.
Raw Water
CO2
2" ID x 60" Plexiglass Columns
Flow Indicator and Control
To Neutralizing Tank
Pump
Mixer
Regen. Sol. Res. 20 gal.

FIGURE 3. Schematic of Acid Mine Drainage Test Apparatus

These chemicals were added in sufficient quantities to achieve the following approximate concentrations.

TABLE 2. Composition of Test AMD Solution.

Free Mineral Acidity (FMA), ppm as $CaCO_3$	500
Sulfate, ppm as SO_4	1150
Calcium, ppm as Ca	200
Magnesium, ppm as Mg	24
Aluminum, ppm as Al	15
Manganese, ppm as Mn	8
Iron, ppm as Fe	210

One of the objectives of this investigation was the determination of the effect of variations of the relative concentrations of ferrous and ferric ions in the synthetic AMD solutions. These variations of concentrations were obtained by varying the relative quantities of $FeSO_4 \cdot 7H_2O$ and $Fe_2(SO_4)_3 \cdot XH_2O$. However, it was intended that the total concentration of both ferrous and ferric iron would always approximate the 210 ppm (as Fe) value.

The regenerant solutions were routinely prepared from commercial chemicals dissolved in demineralized water. Some studies used regenerant solution prepared with either acid mine drainage or ion exchanger column effluents. When this procedure was used, it was specified in the text of this report.

SECTION 6

ION EXCHANGER COLUMN AND LABORATORY OPERATING PROCEDURES

The ion exchange column tests were conducted according to standard operating procedures employed in the ion exchange industry. The following discussion is a review of these procedures so that the reader may understand the cyclic nature of the process. This review also attempts to instruct the reader as to the nomenclature of certain operations which are used repetitively in ion exchange processes and need no further explanation in ensuing discussions. Other definitions can be found in a separate later section.

Ion Exchangers. All the ion exchangers utilized in this study were commercially available from the various manufacturers listed below:

Ion Exchanger	Type	Manufacturer
C-240	Strong Acid Cation Exchanger	Ionac Chemical Sybron Corporation Birmingham, New Jersey 08011
WGR	Weak Base Anion Exchanger (primary amine)	Dow Chemical U. S. A. Dow Chemical Company Midland, Michigan 48640
21K	Strong Base Anion Exchanger Type I - Porous	Ditto
IRA-410	Strong Base Anion Exchanger Type 2	Rohm & Haas Company Philadelphia, Pa. 19105
IRA-68	Weak Base Anion Exchanger (tertiary amine)	Ditto
IRC-84	Weak Acid Cation Exchanger (carboxylic)	Ditto

Ion Exchanger Bed. The sample of particulate ion exchange material in the test columns occupies a given volume within that column because the individual particles settle into a rather compact mass. The compact column of ion exchange material including the interstitial volume, is termed the ion exchanger "bed". Practically all ion exchangers are sold and capacity rated on the basis of the volume occupied by the bed, which has been measured under reproducible conditions.

The ion exchanger bed dimensions are naturally restricted by the inside dimensions of the test column. The test columns used in this study were 2 inches ID by 60 inches high. The height of the ion exchanger bed normally used in this study was 30 inches. Thus, 30 inches of volume above the ion exchanger bed was available for purposes to be revealed below. This volume above the bed is termed the "freeboard".

As stated above, the dimensions of the ion exchanger bed were normally 2 inches in diameter by 30 inches high. This corresponds to an ion exchanger volume of 0.055 cubic feet.

Ion Exchanger "Treatment" Process. The passage of the solution to be treated through the bed of ion exchange material is termed the "treatment step". During the treatment step, the ion exchange capacity is gradually depleted. When the usable capacity is completely used up. the ion exchanger is said to be "exhausted". Therefore, the treatment step is sometimes termed the "exhaustion step".

Treatment flow rates are always expressed in relationship to the volume of ion exchanger used. In this study the treatment flow rate is always expressed in gallons per minute per cubic foot of ion exchanger - gpm/cu ft.

The treatment flows in ion exchange operation may occur either in an upflow or downflow direction. Normal ion exchange processes utilize downflow treatment. However, this investigation involved some studies of upflow treatment processes. Whenever treatment steps are specified, the direction of flow is specified also.

The fluid which is to be treated by the ion exchanger and which is to be introduced into the ion exchanger column in either an upflow or downflow direction is termed the "influent". The treatment fluid which emanates from the ion exchanger column is termed the "effluent".

The treatment step is essentially a batch operation because it continues only as long as the ion exchanger removes the particular ion in question. Therefore a finite volume (batch) of treated water is obtained during the treatment step. The treatment step is normally terminated when the concentration of the contaminating ion in the effluent reaches an undesirable level. This level is usually established beforehand and is based upon the maximum concentration allowable in the product (effluent). The appearance of the contaminating ion in the effluent is sometimes referred to as "leakage".

Ion Exchange Backwash Process. After the treatment step has been completed. a process may be applied to the ion exchanger to restore its capacity for subsequent treatment. Before this process is applied, the ion exchanger bed is usually "backwashed".

Backwashing consists of the passage of water through the ion exchanger column in an upflow direction, causing a fluidization of the bed. This expansion is a result of the particles being borne upward by the flow of water until they reach a zone in the column in which the upward forces of the water match the gravitational pull on the particles. Thus, regulation of backwash flow rates will produce various degrees of expansion of the ion exchanger bed. It is common to apply a sufficient upflow rate to achieve a 50% expansion of the bed. This was the usual practice during this study.

The purpose of the backwash process is to loosen the ion exchanger bed which may have become compacted by the flowage of water through the bed during the treatment step. A second purpose of the backwash process is to remove insoluble materials which are either filtered from the influent during the treatment step or which may be precipitated in the bed. The backwash process was always applied following a treatment step regardless of the direction of flow of the treatment cycle.

Ion Exchange Regeneration Process. The process of restoring the capacity of an ion exchanger by passing a regenerant chemical solution through the ion exchanger bed is termed the "regeneration process" or "regeneration step". The regeneration process may be carried out with various chemicals depending upon the type of ion exchanger involved and the desired form into which the ion exchanger is to be converted. For instance, a strong acid cation exchanger may be regenerated either to the "hydrogen form" by using a strong acid regenerant or the "sodium form" by using sodium chloride as a regenerant. In this study, the regenerant chemical is always specified as is the concentration of the solution used for regeneration.

The flow of the regenerant solution is always related to the volume of the ion exchanger. Thus, the flow will always be specified in this study as gallons per minute per cubic foot of exchanger - gpm/cu ft. The flow may be applied either upflow or downflow and this direction will always be specified.

The actual quantity of chemical regenerant applied to the ion exchanger may be specified in various ways. The common method merely indicates the number of pounds of regenerant applied per cubic foot of exchanger - lbs/cu ft. This data is frequently converted to the number of Kilograins of equivalent calcium carbonate ($CaCO_3$) per cubic foot - Kgrs ($CaCO_3$)/cu ft. This latter value is used to compare the actual capacity obtained by the ion exchanger with the theoretical capacity which would be obtained if regenerant utilization were 100%.

Another method which will be used to express regenerant dosage will be in terms of per cent of the theoretical ion exchange capacity obtained during the previous treatment cycle. This method of expression is used when weak base anion exchangers or weak acid cation exchangers are regenerated. These exchangers are highly efficient in regenerant utilization requiring only a light excess above the theoretical or stoichiometric quantity. In this study, the dosages are expressed as

110% or 100% of the stoichiometric quantity based upon the capacity obtained during the previous exhaustion.

Ion Exchanger Rinse Process. After the regenerant solution has flowed into the ion exchanger column, it is necessary to remove this solution with a rinse fluid, normally in the same flow direction. The rinse fluid serves to displace the last portion of regenerant solution through the ion exchanger bed and also serves to remove the last traces of regenerant from the surface of the ion exchanger particles.

The fluid used to rinse the ion exchanger is usually the same fluid which is to be treated by the ion exchanger. In some of our studies the rinse fluid was demineralized water. The type of rinse fluid used and its flow rate is always specified. The flow rate was always the same flow rate and direction that was used in the regeneration process.

Ion Exchange Capacity. A complete regeneration-exhaustion cycle consists of a regeneration step (backwash, regenerant and rinse) followed by a treatment step. The usual procedure for establishing the capacity of an ion exchanger in the treatment of a given fluid is to perform several complete cycles to reach a "steady state" condition with reproducible results. Two cycles are normally sufficient to reach steady state where the capacity becomes rather constant, within experimental error. In this study, it was the usual practice to ignore data obtained during the initial cycles except to establish that a steady state (stabilized capacity) was attained. Then, the last regeneration-treatment cycle was cited as being typical for the particular set of conditions studied.

Ion exchange capacity was calculated by determining the volume of exhausting AMD water which was passed through the ion exchanger to a predetermined endpoint. The endpoint is that instant when the concentration of contaminant in the treated effluent reaches a specified maximum. As exemplified below, the volume (in gallons) of water treated is multiplied by the concentration (in grains per gallon as $CaCO_3$ - gpg) of the contaminant being removed by the ion exchanger to obtain the number of grains of exchange. This value is then related to the volume of the ion exchanger to obtain grains per cubic foot of exchanger - grs/cu ft. This value usually is in the magnitude of the thousands. Thus, it is more common to express the capacity in terms of Kilograins (1000 grains) per cubic foot - Kgrs/cu ft.

For example, 0.055 cubic feet of ion exchange material treated 20 gallons of AMD to the endpoint. The influent concentration of contaminants removed was 33.6 gpg (as $CaCO_3$).

$$20 \text{ gallons} \times 33.6 \, \frac{\text{grains}}{\text{gals.}} \div 0.055 \text{ cubic feet}$$

equals 12,219 grains per cubic foot; or 12.2 Kilograins per cubic foot.

Aeration and Liming Procedures. Whenever an aeration or liming process was studied, it was carried out batchwise. No attempt was made to measure air flow during aeration. These processes are well known to those versed in the art. Equipment needed to accomplish the processes are predictable from a knowledge of the characteristics of the water to be treated. These processes were carried out on a lab scale only to demonstrate that aeration and lime treatment will produce the desired result.

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SECTION 7

ANALYTICAL PROCEDURES

The analytical procedures used for the analysis of all water solutions are listed below:

pH; Corning Model 7 or Model 10 pH meter with glass/calomel electrodes.

Alkalinity; Titration with 0.02 N H_2SO_4 to a pH of 4.2 with mixed methyl red-bromcresol green indicator endpoint (Standard Methods for the Examination of Water and Wastewater(3), Page 48).

Free Mineral Acidity; Cold titration with 0.02 N NaOH to a pH of 4.5 using methyl orange indicator (Standard Methods(3), Page 46).

Free Mineral Acidity (Hot Titration); Hot titration with 0.02 N NaOH to pH of 8.3 (Standard Methods(3), Page 438-439).

Sulfate; Turbidimetric method using sulfate conditioning solution and barium chloride (Standard Methods(3), Page 291).

Aluminum, Manganese, Sodium, Calcium, Magnesium, Total Iron, Atomic Absorption Spectrophotometry.

Ferrous Iron (Below 10 ppm); O-Phenanthroline Method (Standard Methods(3), Page 156).

Ferrous Iron (10 ppm and over); Titration with 0.0125 N $KMnO_4$ using Ferroin indicator (Ferrous 1-10, phenanthroline) in strong acid solution (modified Zimmerman - Reinhardt method).

The authors feel that it is appropriate at this time to discuss the analytical methods and the relationship of the results obtained to the ion exchange processes which were studied in this investigation.

One very serious problem which was encountered during our studies was the inability to achieve balance with the cation and anion concentrations in the solutions. This was particularly true when analyzing raw AMD solutions or solutions containing large concentrations of iron and aluminum. Although it may be superfluous to discuss the obvious causes for this problem, we feel it is appropriate to do so because an understanding of the problem will help in the interpretation of some of the data.

Because of the difficulty in determining equivalent concentrations of cations and anions in AMD solutions by direct analysis of each ion specie, we believe that the method of determining acidity is yielding a false indication of the hydrogen ion concentration. We contend that the method not only yields a value corresponding to the amount of free acidity in the sample but also includes the acidity resulting from the hydrolysis of acidic salts of ferric iron and aluminum (equations 1, 2 and 3):

$$H_2SO_4 + 2NaOH \longrightarrow Na_2SO_4 + 2H_2O \quad (1)$$

$$Fe_2(SO_4)_3 + 6NaOH \longrightarrow 3Na_2SO_4 + \underline{2Fe(OH)_3} \quad (2)$$

$$Al_2(SO_4)_3 + 6NaOH \longrightarrow 3Na_2SO_4 + \underline{2Al(OH)_3} \quad (3)$$

The analysis is further complicated when the AMD solution contains substantial concentrations of ferrous iron. AMD solutions are not produced and stored under anaerobic conditions. Therefore, some dissolved oxygen will be present in this solution. Although the oxygen will not react with the ferrous iron in the acidic environment, it will react before reaching the 4.2 pH endpoint of the acidity method. Moreover, any atmospheric oxygen introduced during the analytical determination will result in even more acidity production (equation 4):

$$2FeSO_4 + \tfrac{1}{2}O_2 + 4NaOH + H_2O \longrightarrow 2Na_2SO_4 + \underline{2Fe(OH)_3} \quad (4)$$

The determination of acidity by this method may be beneficial as an indicator of the ultimate quantity of a neutralizing agent to be applied to acid mine drainage. However, it is of little value in determining ionic loading factors to cation and anion exchange materials. It is likely that the hot method of determining FMA is even more erroneous even though this method stabilizes the ferrous iron oxidation at its maximum. The error introduced by precipitation of magnesium is a possibility with this method and this prevents accurate back calculations to obtain actual acidity.

Because of these problems and also because of the unreliability of sulfate determinations, we adopted the policy that sulfate concentrations were to be calculated as the difference between the sum of the equivalent cation concentration and the sum of the equivalent chloride and alkalinity obtained by direct analysis. When making this calculation, ferric iron and aluminum were not included with the sum of the equivalent cation concentration because it was assumed that these ions would appear also as free mineral acidity (equations 2 and 3).

The policy was adopted also in situations where FMA was not present (pH above 4.2) because it was assumed that ferric iron and aluminum would be precipitated.

In spite of the procedures described above, some abnormal analytical data will be observed in this report. Most of it is due to the ferrous iron oxidation problem mentioned above.

It is recommended that any future studies involving ion exchange treatment of AMD include a method for the determination of the true acidity concentration. Such a procedure would involve the passage of a sample through a column of a strong acid cation exchanger which is fully converted to the hydrogen form. This converts all metal salts to the corresponding acids which may be determined by titration with sodium hydroxide solutions to a pH of 4.2. Metal ions will not interfere because they are removed. Metal ion determinations should be accomplished by direct methods and converted to calcium carbonate equivalents. The sum of the calcium carbonate equivalents of metal ions subtracted from the acidity of the hydrogen form cation exchanger effluent should be equivalent to the true hydrogen ion concentration.

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SECTION 8

STRONG ACID CATION EXCHANGER PERFORMANCE - HYDROGEN FORM

The ion exchange process which employs a strong acid cation exchanger in the hydrogen form is capable of "splitting" ionized, neutral salts in water. The reaction replaces the cationic portion of the neutral salt with an equivalent hydrogen ion, forming the corresponding acid in the solution phase while the cation is sorbed by the ion exchanger. The following equation (where H-R represents the cation exchanger in the hydrogen form) exemplifies the reaction which takes place during this treatment process:

$$CaSO_4 \quad + \quad \underline{2H\text{-}R} \longrightarrow \underline{Ca = R_2} \quad + \quad H_2SO_4 \qquad (1)$$

The reaction takes place in dilute solutions because of the favorable selectivity of the cation exchanger for multivalent cations.

The reaction can be reversed by the application of more concentrated solutions of strong mineral acids to the cation form of the cation exchanger. The high concentration of hydrogen ions of the regenerant reverses the cation-hydrogen selectivity causing the cation exchanger to release the sorbed cations and return to its hydrogen form.

The process by which the cation exchanger is returned to its original (hydrogen) form is termed the regeneration process. Strong acid cation exchange resins require strong acids for regeneration. Sulfuric acid is normally used because it is much cheaper than other acids. The regeneration is illustrated by the following equation:

$$H_2SO_4 \quad + \quad \underline{Ca = R_2} \longrightarrow CaSO_4 \quad + \quad \underline{2H\text{-}R} \qquad (2)$$

Because high concentrations of acids are required for the regeneration process, the concentration of the sulfate salt (calcium sulfate in equation 2) resulting from the regeneration process will be high also. If a high percentage of calcium ion is sorbed on the ion exchanger, the calcium sulfate produced by the regeneration process could precipitate in the exchanger bed owing to its relatively low solubility. Because of this fact, the usual ion exchange treatment practice is to limit the concentrations of sulfuric acid regenerant to 2% or less or to apply 2% sulfuric acid solutions followed by 5% sulfuric acid solutions in a stepwise fashion when waters containing high concentrations of calcium sulfate are to be treated.

Figure 4 shows a schematic for an AMD treatment process using this cation exchanger to produce a potable product.

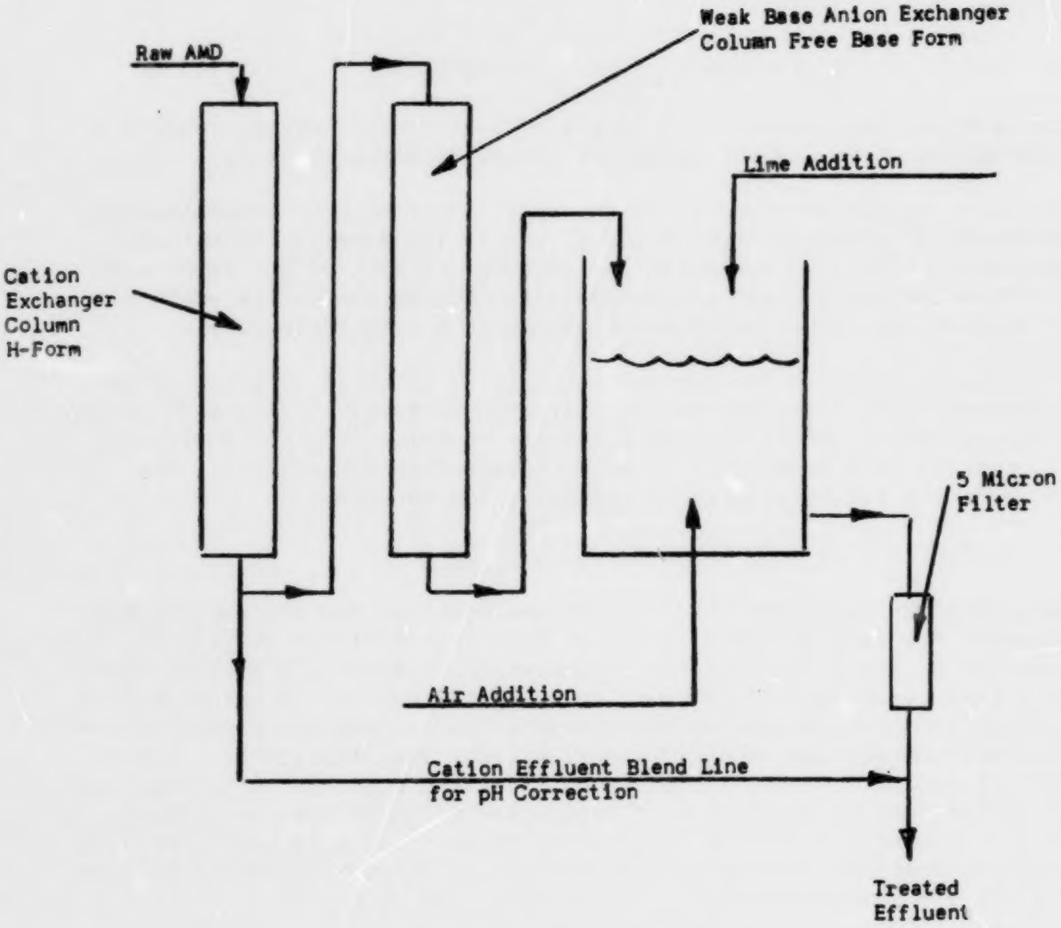
Raw AMD
Weak Base Anion Exchanger Column Free Base Form
Lime Addition
Cation Exchanger Column H-Form
5 Micron Filter
Air Addition
Cation Effluent Blend Line for pH Correction
Treated Effluent

Figure 4 AMD Treatment Process Schematic Hydrogen Form Strong Acid Cation Exchanger/Free Base Form Weak Base Anion Exchanger

Acid mine drainage inherently possesses relatively high concentrations of calcium ion. Thus, the study of this treatment process was concerned with a determination of the characteristics of the treated effluent which would be produced from regeneration procedures which prevented calcium sulfate precipitation in the bed.

Another objective of this study was to determine the effect of variations of sulfuric acid regenerant dosages upon the treated effluent characteristics. This information would establish the minimum quantity of chemical needed to obtain a satisfactory treated water and would be helpful in determining treatment costs.

Our attempts to apply the hydrogen form of a strong acid cation exchange resin to the treatment of AMD have indicated that this process could be employed as the primary step of a complete process for producing potable water. The reason for this conclusion will become apparent from the study which is described below.

All tests were made on an exchanger column measuring 31-33 inches high in the 2" diameter ion exchanger columns. This column represents an exchanger volume of approximately 0.056-0.060 cubic feet. The cation exchanger used in this portion of the study was manufactured by Ionac Chemical Division of the Sybron Corporation and was made with 8% divinyl benzene.

Regenerations of the ion exchanger columns were made with sulfuric acid solutions only. The effect of regenerant dosage, regeneration flow rate and flow direction upon the ion exchange capacity of the resin and the character of the treated water were studied. A "stepwise" regeneration process involving first 2% sulfuric acid followed by 5% sulfuric acid was studied also.

The rinse operation following regeneration was carried out with demineralized water at the same flow rate as was used for regeneration. Demineralized water was used so that the rinse endpoint could be easily recognized.

The backwash step was applied prior to the regeneration step and was carried out with demineralized water at a flow rate sufficient to achieve a 50% expansion of the ion exchanger column.

The AMD treatment step was always carried out in a downflow direction. A study was made of the effect on ion exchanger capacity and treated effluent character when using treatment flow rates of 4 and 8 gpm per cubic foot of ion exchanger.

The effect of sulfuric acid regenerant dosage upon capacity and effluent quality is revealed by the data in Table 3. The treated effluent obtained from the use of a 3 lb/cu ft sulfuric acid regenerant dosage contained relatively high concentrations of iron. At this stage of the investigation it was believed that these high iron concentrations

TABLE 3

Strong Acid Cation Exchanger - Hydrogen Form Performance data.

Run No.	2A	8A	12A	
Sulfuric Acid Dosage, lbs/cu ft	3	6	12	
Regeneration Flow Rate, gpm/cu ft	0.5	0.5	0.5	
Regenerant Flow Direction	Down	Down	Down	
Exhaustion Flow Rate, gpm/cu ft	8.0	4.0	4.0	
Chemical Constituent Effluent				Typical Raw AMD
Ferrous Iron, ppm Fe	25.0	2.8	1.0	204
Ferric Iron, ppm Fe	2.5	2.2	1.6	13
Calcium, ppm Ca	15.0	6.0	3.5	182
Magnesium, ppm Mg	3.3	0.8	0.3	26
Aluminum, ppm Al	N.A.	N.A.	N.A.	18
Manganese, ppm Mn	1.0	0.0	0.0	8.5
Sodium, ppm Na	28.0	0.5	0.0	18
FMA, ppm $CaCO_3$	1430	1540	1575	600
Total Metallic Cations, gpg ($CaCO_3$)	--	--	--	70.0
Ion Exchange Capacity Kilograins/cu ft	10.7	12.7	19.2	--
% Regenerant Utilization	50	30	22	--
Lbs Waste Acid/Kilograins	0.14	0.28	0.49	--

would cause problems during subsequent treatment processes. As an example, if the cation effluent were to be passed through a weak base anion exchanger in the free base form, the iron would precipitate in the ion exchanger bed. The accumulation of these precipitates over a period of time could eventually result in impairment of the efficiency of the ion exchanger.

For this reason, very little time was spent studying the 3 lb/cu ft dosage. In fact, only 2 regeneration/exhaustion cycles were produced and were made at an exhaustion flow rate of 8 gpm/cu ft. However, as will be demonstrated later, exhaustion flow rates between 4 and 8 gpm/cu ft have a negligible effect upon effluent quality. However, a slight increase in capacity and regenerant utilization could be expected at the 4 gpm/cu ft exhaustion rate.

The data in Table 3 shows a marked decrease in metallic cation content of the treated effluent as sulfuric acid regenerant dosage is increased. The ion exchange capacity also increases with increasing

TABLE 4

Strong Acid Cation Exchanger - Hydrogen Form. Performance data

Run No.	12A	19A	12B	19B
Sulfuric Acid Dosage, lbs/cu ft	12	12	12	12
Regenerant Flow Rate, gpm/cu ft	0.5	0.5	1.0	1.0
Regenerant Flow Direction	Down	Down	Up	Up
Exhaustion Flow Rate, gpm/cu ft	4.0	4.0	4.0	4.0
Regenerant Concentration, %	2	2 & 5*	2	2 & 5*
Chemical Constituent Effluent				
Ferrous Iron, ppm Fe	1.0	1.2	3.0	2.0
Ferric Iron, ppm Fe	1.6	1.6	1.1	1.6
Calcium, ppm Ca	3.5	3.5	1.9	2.0
Magnesium, ppm Mg	0.3	0.4	0.3	0.3
Aluminum, ppm Al	N.A.	0.0	0.1	0.0
Manganese, ppm Mn	0.0	0.1	0.2	0.1
Sodium, ppm Na	0.0	0.1	0.1	0.1
FMA, ppm $CaCO_3$	1575	1560	1600	1600
Ion Exchange Capacity, Kilograins/cu ft	19.2	19.2	16.9	19.2
% Regenerant Utilization	22	22	19	22

* 6 lbs/cu ft were applied first at a 2% concentration followed by another 6 lbs/cu ft applied at a 5% concentration.

regenerant dosage as might be expected. However, the above described benefits derived from increased regenerant dosage are obtained at the expense of decreased regenerant utilization. Thus, the amount of acid discharged to waste per Kilograin of exchange capacity appears to increase in direct proportion to the increased acid regenerant dosage. The amount of waste acid by-product discharged by any AMD treatment process is an important consideration both from an economic standpoint and an environmental standpoint.

It appeared at this point that a minimum regenerant dosage of 6 lbs/cu ft was needed to obtain a satisfactory effluent quality for further treatment. It was also desirable to attain the regenerant utilization (50%) achieved by the 3 lb/cu ft dosage. An effort was made to increase regenerant utilization at the 12 lb/cu ft dosage by applying one half the regenerant at a 2% concentration followed by the second half of the regenerant at a 5% concentration. Such stepwise procedures are used in ion exchange practice to prevent calcium sulfate precipitation in the ion exchanger and increase regenerant utilization.

TABLE 5

Strong Acid Cation Exchanger - Hydrogen Form, Performance Data.

Run No.	4A	8A	4B	8B
Sulfuric Acid Dosage, lbs/cu ft	6	6	6	6
Regenerant Flow Rate, gpm/cu ft	0.5	0.5	1.0	1.0
Regenerant Flow Direction	Down	Down	Up	Up
Exhaustion Flow Rate, gpm/cu ft	8	4	8	4
Chemical Constituent Effluent				
Ferrous Iron, ppm Fe	2.8	2.8	2.8	2.1
Ferric Iron, ppm Fe	1.2	2.2	1.0	1.1
Calcium, ppm Ca	6.0	6.0	4.0	2.4
Magnesium, ppm Mg	0.6	0.8	0.7	0.6
Aluminum, ppm Al	N.A.	N.A.	N.A.	N.A.
Manganese, ppm Mn	0.2	0.0	0.2	0
Sodium, ppm Na	4.7	0.5	8.5	0.5
FMA, ppm $CaCO_3$	1650	1540	1450	1510
Ion Exchange Capacity Kilograins/cu ft	11.9	12.7	11.2	12.6
% Regenerant Utilization	28	30	27	30

Results of this effort are illustrated by the data in Table 4. The raw AMD water used as influent for these runs was approximately the same as shown by the typical raw AMD analysis in Table 3. These data show that the stepwise regeneration process was no more efficient than when the regeneration was performed with all the acid at the 2% concentration. The use of a stepwise regeneration increased regenerant utilization when upflow regenerations were carried out. However, the increase merely brought the efficiency of the process up to an equal status with the downflow regeneration process. It must be concluded from these results that stepwise regeneration processes have little value in the treatment of acid mine drainage by hydrogen form cation exchangers.

Table 5 presents data obtained when regeneration direction and exhaustion flow rate were varied using sulfuric acid regenerant dosages of 6 lbs/cu ft. The raw AMD water used as influents was again approximately equivalent to that shown in Table 3. Upflow regenerations produced treated effluents which were slightly lower in metallic ion content. Nearly equivalent capacity and percent regenerant utilization was obtained regardless of the direction of flow at this regenerant dosage.

A flow rate of 8 gpm/cu ft showed a tendency to lower capacity and regenerant utilization. Because it was believed that we should avoid any process which would reduce the percentage of regenerant utilization, it was concluded that regenerations should be conducted downflow at 0.5 gpm/cu ft and treatment flow rates of 4 gpm/cu ft should always be employed.

Tables 60 through 68 (in the Appendix) contain the results of the analysis of effluent samples taken during the course of each run. These data show the variations in concentration of the water constituents during the treatment cycle and show the abrupt increase in metal ion concentration which signals the endpoint of the run. The endpoint throughput gallonage for each run is noted in the tables and was used to calculate the ion exchange capacity set forth in Tables 3 through 5. The total metallic cation content (70.0 gpg as $CaCO_3$) shown in Table 3 for the typical raw AMD solution was used to calculate capacity for all runs.

It should be noted that several runs were made for each set of variables studied. Data presented in this report is representative of a typical "run" for each set of variables. The typical run is usually the last of a series of four or five runs produced to obtain "steady state".

It is apparent from these data that treatment of acid mine drainage with a strong acid cation exchanger in the hydrogen form will remove a substantial quantity of the metal ion impurities. The removal process involves an equivalent exchange of hydrogen ions for the metal ions which are taken up by the cation exchanger. Thus, the treated effluent contained more free mineral acidity than was present in the raw AMD. However, the treated waters having the characteristics shown in Tables 3, 4 and 5 could be subjected to further treatment with a weak base anion exchanger in the free base form to produce a water with characteristics which are nearly within the range of potability. Such a process is illustrated by the following equation:

$$H_2SO_4 + \underline{2RNH_2} \longrightarrow \underline{(RNH_2)_2 \quad \cdot \quad H_2SO_4} \qquad (3)$$

The prospect of being able to produce a potable water using a hydrogen form cation exchanger as part of a complete treatment method led to the decision to include this process with those selected for further study. The results obtained during these preliminary screening tests have indicated that the additional work should include methods which would increase regenerant utilization. Such additional work under this project is discussed in Section 13.

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SECTION 9

STRONG ACID CATION EXCHANGER PERFORMANCE - SODIUM FORM

This process employs a strong acid cation exchange resin. In the sodium state, a cation exchanger would be capable of removing divalent and trivalent metal ions from acid mine drainage according to the following equations:

$$CaSO_4 + \underline{2Na\text{-}R} \longrightarrow \underline{Ca = R_2} + Na_2SO_4 \qquad (1)$$

$$Fe_2(SO_4)_3 + \underline{6Na\text{-}R} \longrightarrow \underline{2Fe = R_3} + 3Na_2SO_4 \qquad (2)$$

This removal process imparts an equivalent quantity of sodium ions to the solution phase as the divalent and trivalent metal ions are absorbed by the cation exchanger.

The regeneration process employs concentrated solutions of sodium chloride to reverse the di- and trivalent ion selectivity of the ion exchanger and return the exchanger to its sodium form. The following equations illustrate the regeneration reaction:

$$\underline{Ca = R_2} + 2NaCl \longrightarrow \underline{2Na\text{-}R} + CaCl_2 \qquad (3)$$

$$\underline{Fe = R_3} + 3NaCl \longrightarrow \underline{3Na\text{-}R} + FeCl_3 \qquad (4)$$

The regeneration step of this treatment process should take place without many difficulties because relatively soluble chloride salts are produced by the regeneration.

Figure 5 shows a schematic for a possible process to treat AMD using the sodium form exchanger followed by other exchangers. The primary objective of the study of this process was to establish the effect of regenerant dosage, exhaustion flow rate and the proportion of ferrous to ferric iron in the AMD upon the character of the treated effluent, the capacity of the ion exchanger and the extent of regenerant utilization.

The results obtained from this study have shown that this process has the capability for the removal of divalent and trivalent cations from AMD. The extent of the removal of these cations is dependent upon the quantity of sodium chloride regenerant applied. Sodium chloride dosages of 15 lbs/cu ft produced treated effluents containing less than 10 ppm of multivalent cations whereas the effluents from regenerations with 8 lbs/cu ft contain approximately 25 ppm of multivalent cations. However, the regenerant utilization at the 15 lbs/cu ft dosage is substantially lower than it is at 8 lbs/cu ft.

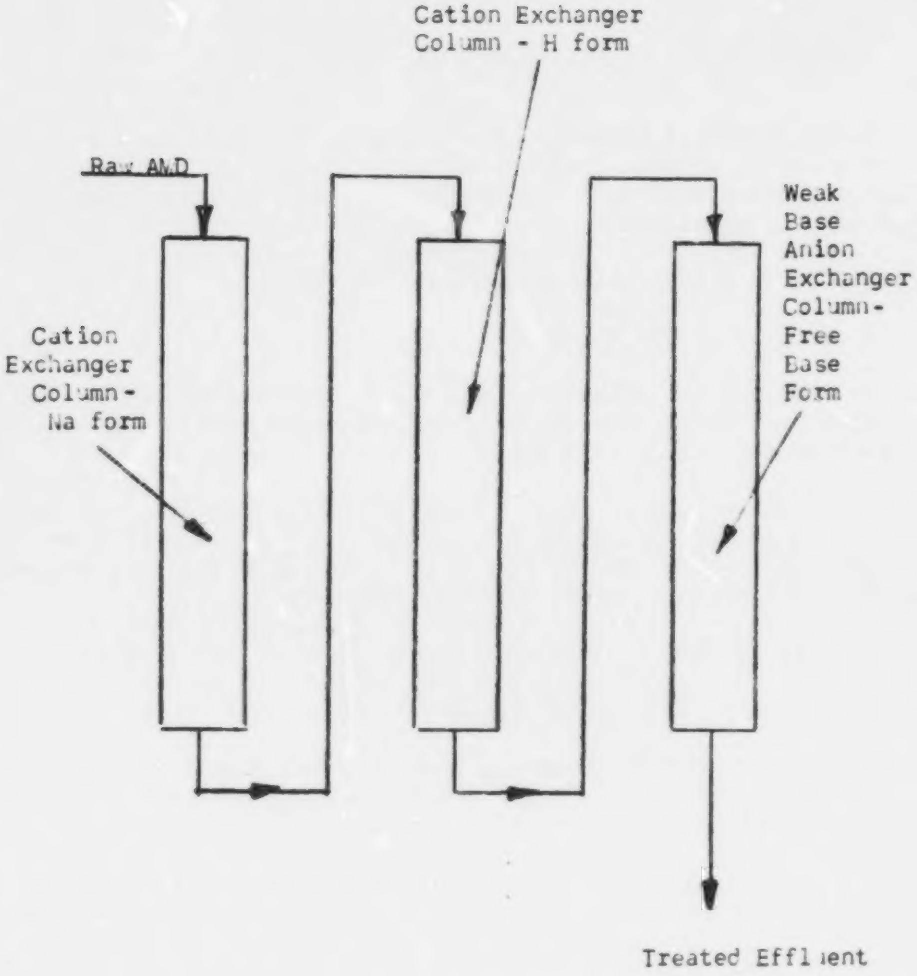
Cation Exchanger Column - H form
Raw AMD
Weak Base Anion Exchanger Column-Free Base Form
Cation Exchanger Column-Na form
Treated Effluent

FIGURE 5. Schematic of Possible AMD Treatment Process Using Sodium Form Cation Exchanger.

The important aspect of this method of treatment is that the treated effluent contains at least the same equivalent of metallic cations that existed in the raw AMD. If the finished water is to meet the requirements for potability, the effluent from this process would require additional treatment to lower the contaminant concentrations to a potability level. A demineralization process would normally be used to accomplish this. In view of the successful results obtained with the hydrogen form of a strong acid cation exchanger, it would be more practical to start with that process because it will remove metallic cations without adding sodium ions to the treated effluent. Subsequent treatment with a weak base anion exchanger in the free base form would lower most of the ionic constituents to the potability level. These conclusions will become apparent from the description of the study set forth below. Therefore, studies on this sodium form resin did not proceed beyond the first phase of the project.

The cation exchanger employed in this study was manufactured by the Ionac Chemical Division of the Sybron Corporation and possessed the same physical and chemical characteristics as the exchanger used in the hydrogen cycle studies.

Regeneration of the ion exchanger was accomplished with 15% (by weight) sodium chloride solutions at a flow rate of 0.5 gpm/cu ft in the downflow direction.

Salt dosages of 8 and 15 pounds of NaCl per cu ft of ion exchanger were used to determine the effect of these variables upon effluent characteristics, ion exchange capacity and regenerant utilization.

After the sodium chloride solution had passed through the ion exchanger, a demineralized water rinse was applied at the same flow rate and direction. The ion exchanger column was backwashed prior to regeneration at a flow rate sufficient to achieve a 50% expansion of the ion exchanger column.

The AMD treatment steps were always carried out in a downflow direction. Treatment flow rates of 4 and 8 gpm/cu ft were employed to determine the effect of this variable upon the quality of the treated effluent, the ion exchange capacity and the extent of regenerant utilization. The effect upon treated effluent characteristics caused by varying the raw AMD ratio of ferrous iron to ferric iron was studied also.

Table 6 shows results obtained by altering salt dosage and treatment flow rate. A comparison of the data for Runs 23B and 23A indicate that the content of iron, calcium and magnesium in the treated effluent obtained by the 15 lbs/cu ft regeneration is much lower than that which is obtained by the 8 lb/cu ft regeneration. Although the ion exchange capacity is lower with the 8 lb/cu ft regeneration than it is with the 15 lb/cu ft regeneration, the per cent of regenerant utilization is higher. These trends were also observed during the study of the strong

TABLE 6

Strong Acid Cation Exchanger - Sodium
Form, Performance Data.

Run No.	23B	23A	26A	
Salt Dosage, lbs/cu ft	8	15	15	
Exhaustion Flow Rate, gpm/cu ft	4	4	8	
Chemical Constituent	Composite Effluent			Typical Raw AMD
Ferrous Iron, ppm Fe	5.8	0.5	1.0	213
Ferric Iron, ppm Fe	1.2	2.5	2.6	3
Calcium, ppm Ca	19.1	5.0	6.6	200
Magnesium, ppm Mg	1.1	0.4	0.5	36
Aluminum, ppm Al	0.1	0.2	0.0	17
Manganese, ppm Mn	0.3	0.1	0.1	7.3
Sodium, ppm Na	475	500	460	2.1
FMA, ppm $CaCO_3$	410	435	445	590
Di & Trivalent metal content, gpg ($CaCO_3$)	--	--	--	66.0
Volume Capacity, gals.	330	450	420	--
Ion Exchange Capacity, grains*	22,000	30,000	28,000	--
Regenerant Utilization, %	46	34	31	--

*Divalent and trivalent metal ion exchange only

acid cation exchanger in the hydrogen form and are a normal occurrence in ion exchange practice. A decision involving the selection of regenerant dosage requires the weighing of various factors including cost of the treatment, the quality of the final water and the ability to remove specific contaminants by subsequent treatment processes. In this case, if the level of sodium as exhibited by these results could be tolerated, the 8 lb/cu ft or an even lower salt dosage should be used to obtain the maximum possible regenerant utilization. Following this, the FMA could be removed by weak base anion exchange and the iron, manganese and aluminum could be removed by liming, aeration and filtration.

Table 6 also compares the results obtained when treatment flow rates of 4 and 8 gpm/cu ft are used. The character of the treated effluent is not materially affected by increasing flow rate from 4 to 8 gpm/cu ft. However, a slight decrease in capacity and percent regenerant utilization was observed.

The free mineral acidity (FMA) content of the composite effluents in Table 6 shows a lower concentration than that of the raw AMD. These data represent the net effect of the behavior of FMA during the course of the treatment cycle. Figure 6 shows the FMA content of the treated effluent during the treatment cycle comparing it to the influent concentration. Note that FMA is almost completely removed during the early stages of the treatment cycle, then gradually increases in concentration and rises above the influent concentration. This means that during the period when the effluent concentration is above the influent concentration, the hydrogen ion sorbed by the cation exchanger at the beginning of the treatment cycle (equation 5);

$$\underline{2Na\text{-}R} + H_2SO_4 \longrightarrow Na_2SO_4 + \underline{2H\text{-}R} \qquad (5)$$

is being displaced by the incoming di- and trivalent metal ions in the raw AMD water (equation 6).

$$\underline{2H\text{-}R} + CaSO_4 \longrightarrow \underline{Ca\text{-}R_2} + H_2SO_4 \qquad (6)$$

The net effect on the composite treated effluent is a modest reduction in FMA content below the raw AMD concentration indicating that some hydrogen form cation exchanger remains after the treatment cycle is terminated. An analysis of the composite regeneration effluent applied after the treatment cycle shows this to be true (Table 7).

The data in Table 8 illustrates the effect of altering the ratio of ferrous to ferric iron in the raw AMD upon the chemical characteristics of the composite treated effluent and the capacity and regenerant utilization of the ion exchanger. These data indicate a negligible effect due to this variable.

Tables 69 through 72 show results of the analysis of effluent samples taken during the course of each run. These data show the variation of effluent ionic constituents as the treatment cycle progresses. The endpoints were characterized by substantial increases in the di- and trivalent cation content of the effluent.

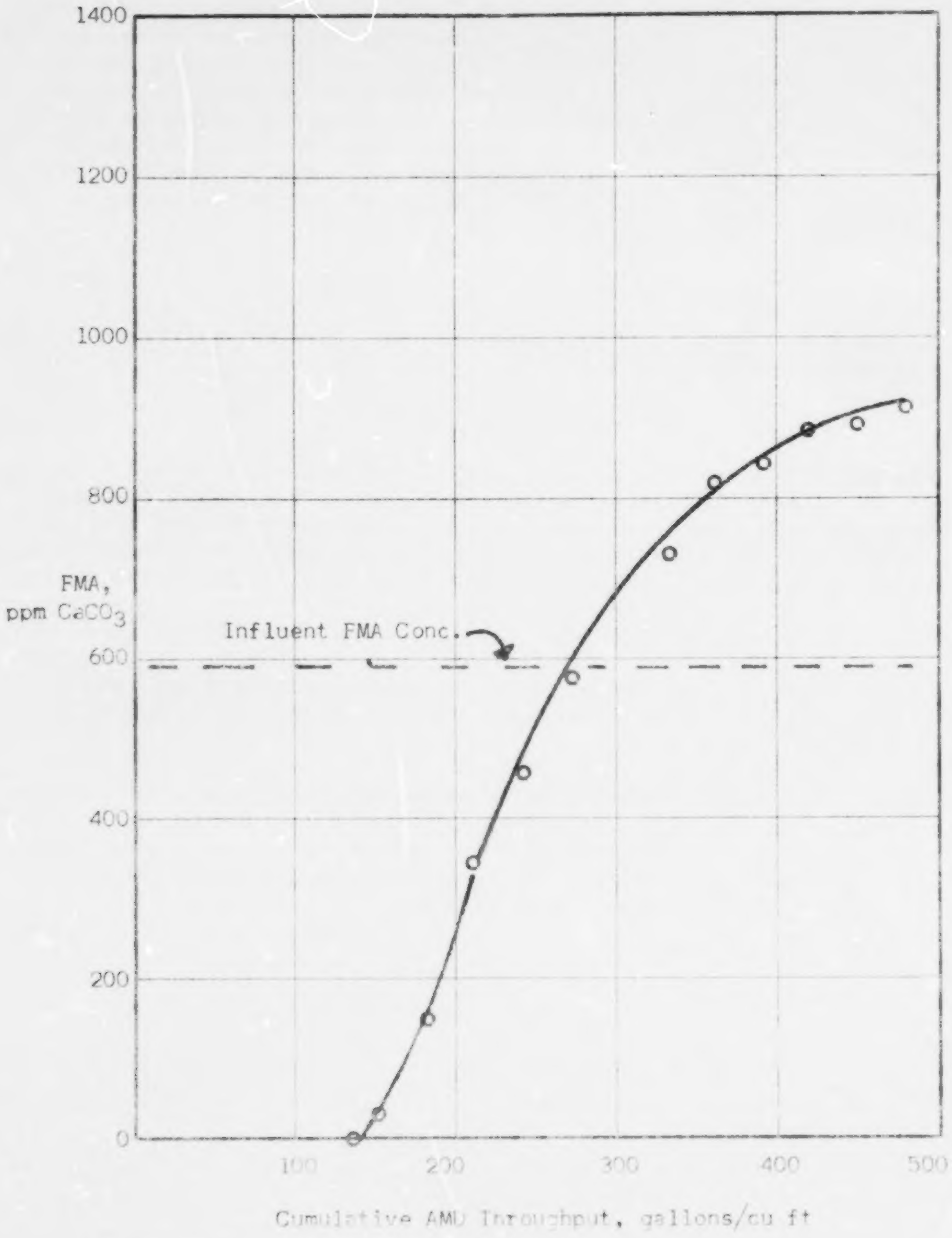

Figure 6 Effluent FMA Concentration vs. Gallons of Acid Mine Drainge Put Through a Sodium State Strong Acid Cation Exchanger.

TABLE 7

Strong Acid Cation Exchanger-Sodium Form. Analysis of Composite Regenerant-Rinse Effluent.

Run No.	23A
pH	1.4
Ferrous Iron, ppm Fe	3,880
Ferric Iron, ppm Fe	220
Calcium, ppm Ca	7,600
Magnesium, ppm Mg	620
Aluminum, ppm Al	330
Manganese, ppm Mn	166
Sodium, ppm Na	17,600
FMA, ppm $CaCO_3$	2,200

TABLE 8

Strong Acid Cation Exchanger - Sodium Form, Performance Data

Run No.	23A	30A
Salt Dosage, lbs/cu ft	15	15
Exhaustion Flow Rate, gpm/cu ft	4	4
Raw AMD Ferrous Iron, ppm Fe	213	140
Raw AMD Ferric Iron, ppm Fe	3	67
Chemical Constituent Effluent		
Ferrous Iron, ppm Fe	0.5	0.6
Ferric Iron, ppm Fe	2.5	2.1
Calcium, ppm Ca	5.0	2.7
Magnesium, ppm Mg	0.4	0.5
Aluminum, ppm Al	0.2	0.1
Manganese, ppm Mn	0.1	0.2
Sodium, ppm Na	500	520
FMA, ppm $CaCO_3$	435	---
Volume Capacity, gals.	450	420
Exchange Capacity, grains	30,000	28,000
Regenerant Utilization, %	34	33

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SECTION 10

WEAK BASE ANION EXCHANGER PERFORMANCE - FREE BASE FORM

This process employs a weak base anion exchange resin. The free base form of this resin is capable of absorbing strong acids. These resins are typically used in multi-column demineralization processes, where strong acids are produced by strong acid cation exchange. However, many acid mine drainage waters contain strong acids without any preliminary treatment. It was the objective of this section of the investigation to study the direct application of acid mine drainage to the free base form of the weak base anion exchanger. We anticipated precipitation of polyvalent cations on the bed, but did not know the real effect on the operating characteristics.

Equation (1) exemplifies the reaction which takes place during the acid removal treatment process.

$$H_2SO_4 + \underline{R\text{-}NH_2} \rightarrow \underline{R\text{-}NH_2 \cdot H_2SO_4} \tag{1}$$

The exhausted form of the weak base anion exchanger as indicated on the right side of equation (1) is termed the "sulfate salt form" or "salt form" of the weak base anion exchanger.

The most important consideration of the treatment of acid mine drainage by this process is the fate of certain metallic elements (ferric iron and aluminum) which are also present in the acidic AMD solution but which will precipitate when the acidity is removed. Equation (2) shows how this can occur:

$$Fe_2(SO_4)_3 + 6H_2O + \underline{3R\text{-}NH_2} \rightarrow \underline{2Fe(OH)_3} + \underline{3R\text{-}NH_2 \cdot H_2SO_4} \tag{2}$$

It is a certainty that iron and aluminum hydroxides will precipitate in the weak base anion exchanger during the treatment cycle. Because there is little chance that the backwash or regeneration cycle will remove these precipitates completely, (except for the aluminum specie) one of the objectives of this study was to determine if the accumulation was detrimental to the operation of the exchanger. Another objective was to determine if the rate of accumulation could be slowed by carrying out the treatment process in an upflow direction. Such a procedure would not allow as much insoluble material to be "filtered out" by the particular ion exchanger bed.

The regeneration process consists of the application of a solution of ammonium hydroxide or an alkali solution to the salt form of the weak base anion exchanger. Equation (3) illustrates this reaction.

$$\underline{R\text{-}NH_2} \cdot \underline{H_2SO_4} + 2NaOH \rightarrow \underline{R\text{-}NH_2} + Na_2SO_4 + 2H_2O \tag{3}$$

This reaction takes place because the resulting neutral salt (Na_2SO_4) cannot be adsorbed by a weak base anion exchanger. Weak base anion exchangers do not have the ability to "split" neutral salts. Because the regeneration process is essentially an acid/base neutralization reaction, the regenerant utilization will approach 100%. It is customary in ion exchange practice to apply a regenerant dosage of 110% of the stoichiometric quantity of acid adsorbed during the previous treatment cycle when using weak base anion exchangers.

Our attempts to apply weak base anion exchangers to the direct treatment of acid mine drainage have been successful. Although we were not able to prevent rapid accumulation of iron hydroxides on the weak base anion exchanger particles, the presence of these contaminants did not appear to impair the efficiency of the ion exchange process. This led us to the conclusion that any cation exchange process which might precede the weak base anion exchange process could be operated at low, more efficient regenerant dosages. This process was one of those selected for further study in the second phase of the project.

Figure 4 is a schematic for an AMD treatment process using this exchanger to produce a potable product. This process was established as being feasible during this project. Initial studies with this weak base anion exchanger considered using the exchanger in a three exchanger system: As a "roughing exchanger" to remove the majority of natural acidity in AMD, followed by the process shown in Figure 5.

The weak base anion exchanger which was used in this series of tests was manufactured by the Dow Chemical Company and is designated by them as WGR.

All tests were made on an exchanger column measuring 31-33 inches high in the 2" diameter ion exchanger columns. This represents an exchanger volume of approximately 0.056-0.060 cu ft.

Regeneration of the columns was accomplished with 4% sodium hydroxide solutions at a flow rate of 0.44 gpm/cu ft downflow. The dosage of sodium hydroxide was always 110% of the stoichiometric quantity of acid adsorbed during the previous treatment cycle.

The rinse operation following regeneration was carried out with demineralized water at the same flow rate as was used for regeneration.

The backwash step applied prior to the regeneration step was performed using demineralized water at a flow rate sufficient to achieve a 50% expansion of the ion exchanger column.

As stated above, the study of this process emphasized methods of retarding the accumulation of iron precipitates in the ion exchanger bed. Upflow treatment cycles were compared to downflow treatment. A downflow treatment flow rate of 4 gpm/cu ft is compared to the 2 gpm/cu ft down-

TABLE 9

Weak Base Anion Exchanger - Free Base Form
Performance Data

Run No.	39B	39A	34B	
Treatment Flow Rate, gpm/cu ft	4	2	2	
Treatment Flow Direction	Downflow	Downflow	Upflow	
Chemical Constituent	Composite Effluent			Typical Raw AMD
Ferrous Iron, ppm Fe	77	58	66	109
Ferric Iron, ppm Fe	6.3	3.0	22	113
Calcium, ppm Ca	200	200	200	200
Magnesium, ppm Mg	30	30	30	30
Aluminum, ppm Al	8	14	8	16
Manganese, ppm Mn	7.8	7.0	8.0	7.9
Sulfate, ppm SO_4	500	480	493	1197
Chloride, ppm Cl	195	184	183	206
pH	4.5	5.0	5.1	2.45
FMA, ppm $CaCO_3$	0	0	0	555
FMA, gpg $CaCO_3$	--	--	--	32.5
Volume Capacity, gals/cu ft	720	970	850	--
Ion Exchange Capacity, grains/cu ft	23,400	31,500	27,500	--

TABLE 10

Weak Base Anion Exchanger
Iron Content Analyses

Sample No.	A	B
Iron content before cycling, %	0.037	0.037
No. of downflow cycles	10	6
No. of upflow cycles	0	4
Iron content after cycling, %	38.3	32.7
Iron content after iron removal treatment, %	0.09	0.07

flow treatment also. Table 9 compares the data obtained by these variations.

The chemical characteristics of the composite effluent obtained by the 4 gpm/cu ft treatment flow rate (downflow) is very similar to that obtained by the 2 gpm/cu ft flow rate (downflow). The ferrous iron concentration is somewhat higher. But, this could be explained by the slightly lower pH. The important consideration is the lower ion exchange capacity which results from the higher treatment rate. The lower capacity does not affect regenerant utilization in this process because regenerant dosages are always 110% of the stoichiometric amount required (91% regenerant utilization).

It is apparent that a 4 gpm/cu ft treatment flow rate could be used for the direct treatment of AMD solutions having the free mineral acidity content exhibited in Table 9. However, it was considered unlikely that this flow rate could be used to treat hydrogen form cation effluents with roughly three times the concentration of FMA.

Table 9 also shows the data obtained by an upflow treatment cycle at 2 gpm/cu ft. The ferrous iron concentrations produced by the upflow and downflow runs at the same flow rate are similar. However, there is a substantial difference in the ferric iron concentrations. Because ferric iron is considered the insoluble specie, while ferrous iron is the soluble specie at this pH, it appears that the upflow exhaustion is permitting more insoluble material to appear in the treated effluent. Thus this method could be useful in retarding the accumulation of insoluble material in the ion exchanger bed when the direct treatment of AMD by this process is attempted. However, it was decided that the upflow process was not needed when the weak base anion exchanger was required to treat a hydrogen form cation exchanger effluent. This decision was based upon the iron concentrations being much lower in the cation effluent than they were in the test water. In turn, the iron retained by the bed would be much smaller also—perhaps of less concern than the lower capacity obtained by the upflow operation.

The weak base anion exchanger in each of the two columns was analyzed for iron content after 10 exhaustion cycles were obtained. Table 10 shows results of the iron analysis made upon the ion exchanger samples before cycling, after cycling and after iron removal treatments on the weak base anion exchanger were attempted.

The iron accumulation in the anion exchange material caused a volume increase of about 20% above the original volume. This result indicates that some method of removing iron from the ion exchanger may be required at periodic intervals. The iron removal treatment employed to obtain the result shown in Table 10 consisted of soaking the ion exchange material in 10% hydrochloric acid at 150°F for 2 hours. Three such treatments were required for these samples suggesting that iron removal treatments should be applied before the iron content reaches the level attained in this study.

The interesting aspect of this process was the fact that although massive accumulations of iron on the ion exchange particles were realized, there was no apparent change in the ion exchange capacity or the kinetics of the ion exchange material. It would have been reasonable to expect the coating of iron oxide on the ion exchanger particles to slow the diffusion rate of acid through the coating or to block ion exchange sites of the ion exchanger. Apparently, the iron oxide (hydroxide) is sufficiently porous to allow sulfuric acid to diffuse freely.

Tables 73, 74 and 75 are presented to show the character of the effluent during the course of the treatment cycles. The effluent analysis for runs 39A and 39B indicate that very little alkalinity is imparted to the treated effluent during the run. This was the case with most of the runs which were made. However, a few runs produced results exemplified in Table 74. During the initial portion of these runs, a significant concentration of alkalinity was imparted to the treated water. This suggests that the WGR weak base anion exchanger may be capable of existing in the bicarbonate form in the same manner that IRA-68 does. (See Chapter 11).

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SECTION 11

WEAK BASE ANION EXCHANGER PERFORMANCE - BICARBONATE FORM

This process employs the unusual characteristics of a unique weak base anion exchange resin, Amberlite IRA-68, manufactured by Rohm and Haas. In the free base form, this resin (represented as R-NH in equation 1 below) is capable of adsorbing carbonic acid to form the bicarbonate salt. This bicarbonate salt has a bicarbonate/sulfate, chloride selectivity such that neutral salts in water can be converted to bicarbonate salts. For example, the sulfate salt of calcium or magnesium is converted to the corresponding bicarbonate salt as illustrated by equation (1).

$$\underline{2(R\text{-}NH_2)HCO_3} + CaSO_4 \longrightarrow \underline{(R\text{-}NH_2)_2SO_4} + Ca(HCO_3)_2 \tag{1}$$

During the treatment of acid mine drainage, FMA would be removed by the reaction illustrated by equation (2).

$$\underline{2(R\text{-}NH_2)HCO_3} + H_2SO_4 \longrightarrow \underline{(R\text{-}NH_2)_2SO_4} + 2H_2CO_3 \tag{2}$$

The effluent from this process would consist primarily of bicarbonate salts of calcium, magnesium, ferrous iron and manganese. In the resultant alkaline environment, it is reasonable to expect ferric iron and aluminum to be precipitated. This precipitation will occur in the exchange bed, or in the effluent liquid. The effluent from the exchanger can be aerated and lime treated to obtain a potable effluent as illustrated by equations (3), (4), (5) and (6).

$$2Fe(HCO_3)_2 + \tfrac{1}{2}O_2 + H_2O \longrightarrow \underline{2Fe(OH)_3} + 4CO_2 \tag{3}$$

$$Ca(HCO_3)_2 + Ca(OH)_2 \longrightarrow \underline{2CaCO_3} + 2H_2O \tag{4}$$

$$Mg(HCO_3)_2 + 2Ca(OH)_2 \longrightarrow \underline{2CaCO_3} + \underline{Mg(OH)_2} + 2H_2O \tag{5}$$

$$Mn(HCO_3)_2 + \tfrac{1}{2}O_2 + 2Ca(OH)_2 \longrightarrow \underline{MnO_2} + \underline{2CaCO_3} + 3H_2O \tag{6}$$

Regeneration of the weak base anion exchanger is accomplished by first contacting the exchanger with a sodium hydroxide solution to convert the ion exchanger to the hydroxyl form (Equation 7).

$$\underline{(R\text{-}NH_2)_2SO_4} + 2NaOH \longrightarrow \underline{2R\text{-}NH_2OH} + Na_2SO_4 \tag{7}$$

Following a rinse operation to remove excess sodium hydroxide, the anion exchange resin is contacted with a solution containing carbon dioxide to convert the exchanger to the bicarbonate form (Equation 8).

$$\underline{R\text{-}NH_2OH} + H_2CO_3 \longrightarrow \underline{(R\text{-}NH_2)HCO_3} + H_2O \tag{8}$$

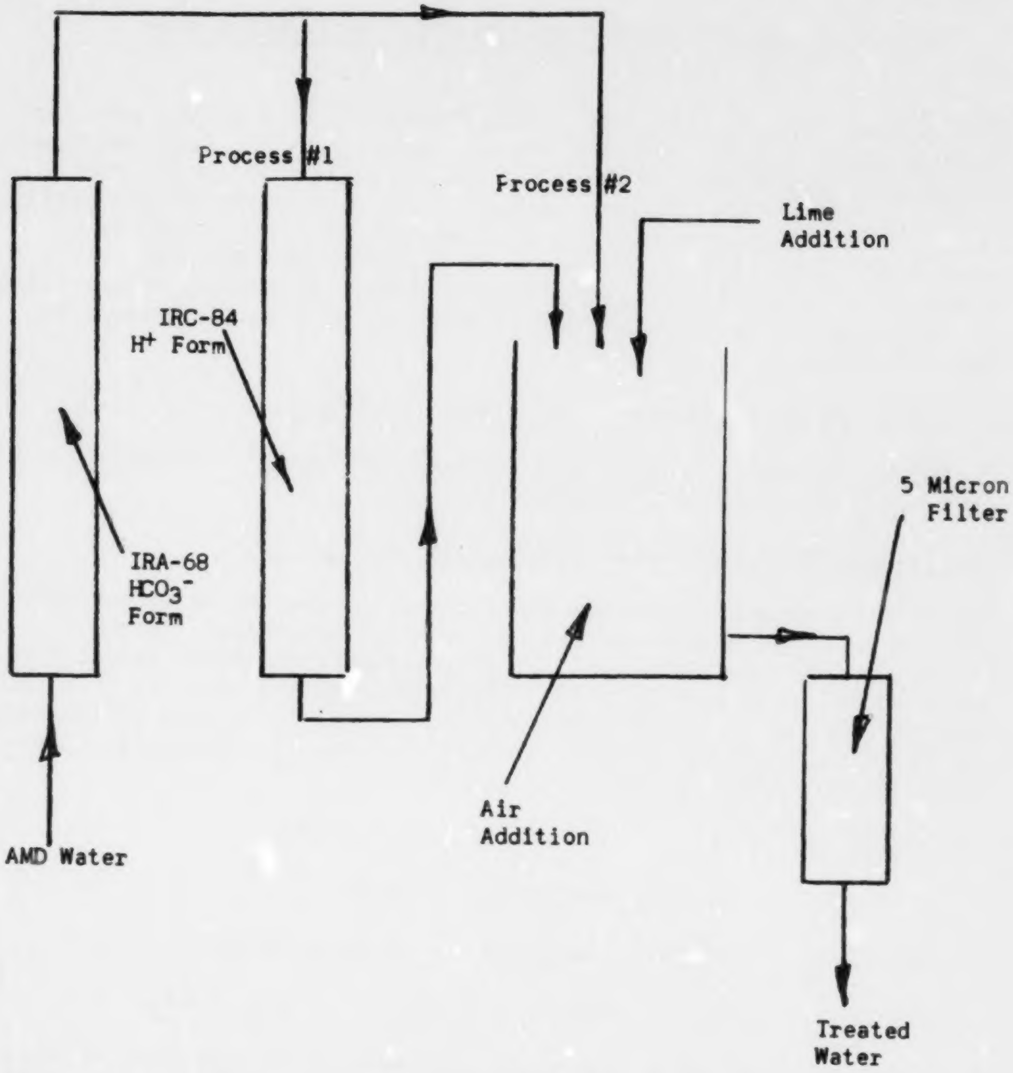
Process #1
Process #2
Lime Addition
IRC-84 H+ Form
IRA-68 HCO3- Form
5 Micron Filter
Air Addition
AMD Water
Treated Water

Figure 7 AMD Treatment Process Schematic Bicarbonate Form Weak Base Anion Exchanger/Hydrogen Form Weak Acid Cation Exchanger or Lime Treatment

Figure 7 is a schematic of an AMD treatment process to use this exchanger to produce a potable product.

Our initial attempt in the first phase of the project to apply the bicarbonate form of a weak base anion exchange resin to the treatment of acid mine drainage was successful. This treatment method removed FMA and converted approximately 60% of the remaining neutral salts to bicarbonate salts. Because of these successes, this process was included among those selected for further study in phase two of this project (see Section 15).

Regeneration of the ion exchanger was accomplished by backwashing briefly, followed by passing a 4% sodium hydroxide solution through the bed at a flow rate of 0.45 gpm in a downflow direction. The amount of caustic applied was 110% of the theoretical quantity of the total FMA removed from the AMD plus that amount of alkalinity imparted to the treated water. The excess caustic was removed by rinsing with demineralized water at the same flow rate. The resin was then transformed to the bicarbonate form by passing a saturated solution of carbon dioxide under 50-60 psig pressure upflow through the bed at 0.5 gpm/cu ft until the effluent pH was 4.5. No rinse was applied after the carbonation step.

Initially, downflow exhaustions at 2 gpm/cu ft were attempted for the treatment cycle of this process. Downflow exhaustions were abandoned early because precipitates formed in the ion exchanger bed and caused flow stoppages during the treatment cycle. Upflow treatment cycles at a flow rate of 2 gpm/cu ft were used during the remainder of the study and no problems were encountered with this method of operation. No other variables were studied other than the attempts at upflow and downflow treatment cycles.

TABLE 11

Weak Base Anion Exchanger - Bicarbonate Form.
Water Analyses of influent and composite effluent.

Chemical Constituent	Composite Effluent	Raw AMD
Ferrous Iron, ppm Fe	48	64
Ferric Iron, ppm Fe	39	136
Calcium, ppm Ca	190	190
Magnesium, ppm Mg	29	30
Aluminum, ppm Al	9	16
Manganese, ppm Mn	7.4	7.4
Sulfate, ppm SO_4	23	1,006
Chloride, ppm Cl	114	219
pH	5.7	2.45
Alkalinity, ppm $CaCO_3$	510	0
FMA, ppm $CaCO_3$	0	600
Volume Capacity, gals/cu ft	485	--
Ion Exchange Capacity, grains/cu ft	31,500	--

TABLE 12
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 44B (Wk. Base HCO_3^-)

Date: 12/14/70

Run No.										
Throughput, gals/cu ft	62	190	380	445	475	485	490			
Ferrous, ppm Fe	34	34	39	48	53	53	56			
Ferric, ppm Fe	45	24	35	65	87	97	94			
Calcium, ppm Ca	186	185	186	195	195	196	189			
Magnesium, ppm Mg	29	29	29	28	28	28	28			
Aluminum, ppm Al	5.0	2.6	6.6	9.8	11	12	13			
Manganese, ppm Mn	7.4	7.4	7.9	8.2	8.1	8.0	7.4			
Sodium, ppm Na	1.7	2.1	0.9	1.3	0.7	0.7	0.7			
FMA, ppm $CaCO_3$	0	0	0	0	0	20	100			
Hot FMA, ppm $CaCO_3$					210	300	380			
Sulfate, ppm SO_4	0	0	0	295	390	390	370			
Chloride, ppm Cl	0	0	68	190	213	222	222			
Alkalinity, ppm $CaCO_3$	685	690	590	130	5	0	0			
pH	6.1	6.3	6.6	5.7	4.4	3.3	3.0			
Sp Resistance, ohm-cm										
Temperature, °F	73	74	74	74	74	74	74			
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

Table 11 shows analyses of the influent AMD and composite effluent obtained on the fourth regeneration/exhaustion cycle using an upflow exhaustion rate of 2 gallons per minute per cubic foot of exchanger. Table 12 shows the characteristics of the treated effluent during the course of the run.

These data show a large decrease in ferric iron concentration from the influent concentration. Because very little iron appears in the regeneration effluent and backwash effluent, a significant amount of iron must be retained by the ion exchanger particles. The weak base anion exchanger was analyzed after four complete regeneration/exhaustion cycles and was found to contain 1.25% Fe based upon 105°C dried free base form resin. No apparent capacity losses were observed over the four cycles. However, this number of cycles may be insufficient to establish a trend. Additional work is necessary to study this affect.

Table 11 shows that metal ion concentrations other than iron and aluminum are relatively unchanged by the treatment process. Sulfate ion is about 98% removed, while chloride ion is about 50% removed. It should be mentioned here again that chloride ion is normally absent in acid mine drainage. It was present in the synthetic AMD used here because of the unavailability of ferric sulfate reagent. This situation was corrected in the later studies of this process (see Section 15).

The treated effluent from this process is shown to be free of FMA, while a large concentration of alkalinity in the form of bicarbonate ion has been imparted to it. This indicates that further treatment with lime should result in an acceptable water from a potability standpoint.

The ion exchange capacity shown in Table 11 was obtained by multiplying the volume capacity (in gallons) by a loading factor. This factor was obtained by adding the concentration of FMA (in grains per gallon as $CaCO_3$) removed from the AMD to the concentration of alkalinity (also in grains per gallon as $CaCO_3$) created in the composite treated effluent. This capacity agrees with the capacity claimed by the manufacturer of IRA-68.

TABLE 13

Weak Base Anion Exchanger - Bicarbonate Form.
Analysis of Composite Regenerant Effluent.

pH	6.6
Ferrous Iron, ppm Fe	3.0
Ferric Iron, ppm Fe	6.5
Calcium, ppm Ca	16
Magnesium, ppm Mg	0.4
Aluminum, ppm Al	26
Manganese, ppm Mn	0.2
Sodium, ppm Na	5400
Alkalinity, ppm $CaCO_3$	425
Sulfate, ppm SO_4	9650
Chloride, ppm Cl	1160

Table 13 illustrates the characteristics of the combined sodium hydroxide regeneration and rinse effluent obtained after a treatment cycle. These data indicate that the major constituents of the waste effluent are sodium sulfate, sodium chloride (normally absent in natural AMD) and sodium bicarbonate. The excess sodium hydroxide appears to be converted to sodium bicarbonate by the residual bicarbonate ion remaining on the anion exchange sites after the treatment step is ended.

Results obtained by this study indicate that this treatment process could be capable of producing a potable effluent. Thus, studies with this resin were continued in phase two of this project, as discussed in Section 15.

SECTION 12

STRONG BASE ANION EXCHANGER PERFORMANCE - SULFATE FORM

The process utilizes the ability of conventional strong base anion exchangers to operate in a so-called sulfate-bisulfate cycle. This cycle is analogous to the second dissociation of sulfuric acid in water (equation 1):

$$HSO_4^- \rightleftharpoons H^+ + SO_4^= \qquad (1)$$

In the presence of excess acid (H^+), this equilibrium is shifted to give increased concentrations of HSO_4^-. In the resin, with non-acidic solutions, the divalent sulfate ion occupies two exchange sites. When this form of resin is contacted by acidic solutions, the monovalent bisulfate ion is formed freeing one resin site which then may be occupied by another anion (equation 2):

$$\underline{R = SO}_4 + H_2SO_4 \rightleftharpoons \underline{R = (HSO}_4)_2 \qquad (2)$$

The regeneration of the anion exchange resin is essentially a reversal of the treatment step, i.e., the conversion of the resin from the bisulfate form back to the sulfate form. A shift of the equilibrium in favor of the sulfate ion can best be obtained by a lowering of the acidity in the anion bed. This lowering of acidity can be accomplished by rinsing the anion exchanger with lime treated AMD water. The addition of lime to AMD neutralizes its free mineral acidity and results in calcium sulfate formation. The resulting solution is separated from the insoluble material and is passed into the ion exchanger column. The elevated pH of this solution causes the sulfate-bisulfate equilibrium to shift in favor of sulfate ion (equation 3):

$$\underline{R = (HSO}_4)_2 \longrightarrow \underline{R = SO}_4 + H_2SO_4 \qquad (3)$$

Theoretically, it is not necessary to neutralize the sulfuric acid formed in equation 3 because this acid will contact resin sites already in the bisulfate form as flow passes down through the ion exchanger bed. However, the reaction rate of equation 3 may be increased by the addition of alkali in the rinse solution applied to the anion exchanger.

The attempt to apply this process to the direct treatment of acid mine drainage has been only partially successful. The treatment process has resulted in complete removal of FMA from the AMD. However, we were not able to obtain predicted FMA removal capacity from regenerations performed with limed AMD waters. Predicted FMA capacity was approached by the use of 2% ammonium hydroxide regenerating solutions. Because of this result and the fact that the process was an inherently low capacity, this process was not recommended for further study beyond phase one of this project. A description of the study follows.

The anion exchangers used during this study were manufactured by the Dow Chemical Company (exchanger designated as 21K) and the Rohm and Haas Company (exchanger designated as IRA-410). The 21K exchanger is considered a type 1 porous strong base anion exchanger and the IRA-410 is considered a type 2 non-porous strong base anion exchanger.

All tests were performed on columns measuring 2 inches in diameter by 30 inches in height (ion exchanger volume equals 0.055 cubic feet).

Regenerations were accomplished by backwashing and then by passing either limed AMD water or a 2% ammonium hydroxide solution through the exchanger column in a downflow direction.

When limed AMD regenerations were performed, the limed solution was passed through the exchanger at a flow rate of 3.0 gpm/cu ft until the effluent pH was equal to the influent pH. This period of time varied between 60 and 75 minutes. The column was then rinsed with deionized water. Limed AMD solutions were prepared by adding lime to synthetic AMD water until the pH was 8.5-8.8. Then the resulting precipitates were allowed to settle prior to decanting and filtering of the supernatant.

When the NH_4OH regenerations were performed, a quantity of 2% NH_4OH solution corresponding to 100% of the stoichiometric quantity of acid removed during the previous exhaustion was passed through the exchanger at a flow rate of 0.25 gpm/cu ft. The column was then rinsed with demineralized water at the same flow rate.

In all cases, exhaustions were carried out downflow at a flow rate of 2 gpm/cu ft. Exhaustions were performed with synthetic AMD waters having FMA concentrations at two different levels to study the effect of this variable upon the capacity of the exchanger.

Table 14 compares the volume capacity and ion exchange capacity obtained from the two anion exchangers when regenerated with limed AMD and with 2% ammonium hydroxide. Table 14 also compares the capacity obtained when the concentration of FMA in the acid mine drainage was varied.

The capacities derived from limed AMD regenerations are shown to be considerably inferior to those obtained using ammonium hydroxide regenerations. Liming AMD to a higher pH may produce higher capacities. However, the prospects for obtaining a practical capacity using limed AMD appeared so remote that it was elected to abandon this portion of the study and concentrate on ammonia regenerations.

Table 14 shows a prominent superiority of the type 2 non-porous exchanger (IRA-410) over the type 1 porous exchanger (21K) in acid removal capacity at either FMA concentration. Table 14 also indicates a modest (15%) increase in acid removal capacity exhibited by the 21K exchanger when the influent FMA was increased from 535 ppm to 860 ppm. However, a

TABLE 14

COMPARISON OF ACID REMOVAL CAPACITY OF SULFATE FORM STRONG BASE ANION EXCHANGER

Run No.	Resin	Regenerant	Influent FMA ppm $CaCO_3$	Volume Capacity gallons/cu ft	Acid Removal Capacity grains*/cu ft
49A	21K	Limed AMD	535	27	840
50A	IRA-410	Limed AMD	540	21	660
49B	21K	2% NH_4OH	535	49	1530
55B	21K	2% NH_4OH	860	35	1760
52A	IRA-410	2% NH_4OH	545	70	2130
55A	IRA-410	2% NH_4OH	860	58	2920

*As $CaCO_3$

TABLE 15

Typical Influent and Effluent Analysis For a Sulfate Form Strong Base Anion Exchanger

Run No.	52A	
Chemical Constituent	Effluent	Raw AMD
Ferrous Iron, ppm Fe	62	104
Ferric Iron, ppm Fe	2	96
Calcium, ppm Ca	198	201
Magnesium, ppm Mg	27	28
Aluminum, ppm Al	13	15
Manganese, ppm Mn	7.9	7.9
Sulfate, ppm SO_4	462	1048
Chloride, ppm Cl	102	196
pH	5.2	2.40
FMA, ppm $CaCO_3$	--	545
Alkalinity, ppm $CaCO_3$	118	--

substantial increase (37%) was exhibited by the IRA-410 with the same influent FMA increase. If this process were to be considered for AMD treatment, the type 2 anion exchanger would be selected for use.

Table 15 displays the character of the composite effluent obtained by this treatment method when the IRA-410 resin was regenerated with ammonia. The process removed nearly 100% of the ferric iron and substantial quantities of ferrous iron. Ferric iron was undoubtedly precipitated and filtered in the ion exchanger bed. Some of the ferrous iron was probably oxidized because of the aerobic conditions under which the AMD was prepared and met the same fate as the ferric iron in the exchanger bed.

Note that a relatively large concentration of alkalinity was produced in the effluent. It was likely that the strong 2% ammonia solutions were capable of converting some of the anion exchange sites to the hydroxyl form. Weaker solutions of ammonia probably would not have this capability and may produce higher acid absorption capacities. Tables 76 through 81 are presented to show the character of the treated effluent during the course of the treatment cycle. These data show that a large amount of ferric iron is removed by this process. The ferric iron is precipitated and becomes deposited upon the ion exchanger as the treatment process progresses. The backwash operation and the regeneration operation do not remove all of this precipitated iron. Thus, an accumulation is expected requiring eventual removal treatments.

Based upon the results obtained above, this process was not selected for further study. This decision was based upon the success of other processes which offer a greater chance for success in producing a potable effluent from acid mine drainage.

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SECTION 13

COMPLETE PROCESS EVALUATION
STRONG ACID CATION EXCHANGER (HYDROGEN FORM)

As stated earlier (Section 4), the second objective of this project was to evaluate three of the most promising treatment processes selected from the original five processes. The process utilizing a strong acid cation exchanger in the hydrogen form was one of the three selected for further evaluation.

The hydrogen form cation exchanger process is not a complete treatment method. However, it will be incorporated as the primary treatment step of a complete process utilizing a weak base anion exchanger in the free base form (Section 14). The purpose of the investigations covered by this section was to determine the effects of applying this treatment method to acid mine drainage containing iron either entirely in the ferrous state or entirely in the ferric state. Another objective of this investigation was the determination of whether regenerant utilization could be increased by recovering and reusing waste sulfuric acid regenerant. Still another objective of this study was an investigation of the economics of using an alternate regenerant (hydrochloric acid) in comparison to that obtained by using sulfuric acid.

The results of these investigations, as detailed below, have indicated that the waste sulfuric acid or hydrochloric acid regenerant is not suitable for reuse. The study has also shown that cation regeneration chemical costs would be approximately 45-60% higher if hydrochloric acid were used rather than sulfuric acid. The effect on this treatment process of the type of iron (ferrous or ferric) in the synthetic AMD on the characteristics of the treated effluent were negligible. However, better iron removal was obtained when the iron was present entirely in the ferric form.

Reactions involved with this treatment process are shown in Section 8 and the manipulation of the ion exchanger columns was carried out as described in that section.

The first phase of this investigation used AMD solutions containing iron essentially in the ferrous state. This phase emphasized the use of hydrochloric acid as a regenerant and attempts were also made at waste hydrochloric acid regenerant recovery and reuse.

Table 16 compares the results obtained by varying hydrochloric acid dosage. Variations of regenerant concentration and flow rate were studied also and the effect of these variables is demonstrated by comparing the data from Runs No. 69A and 74A. The data indicates that variations of regenerant concentration between 4.6% and 9.2% have little affect upon the capacity of the cation exchanger or the character of the

TABLE 16.

Strong Acid Cation Exchanger - Hydrogen Form.
Performance Data.

Run No.	64A	69A	74A	77B
Regenerant	HCl	HCl	HCl	HCl
Regenerant Dosage, lbs*/cu ft	6	4	4	2
Regenerant Dosage, Kgrs($CaCO_3$)/cu ft	57.6	38.4	38.4	19.2
Regenerant Conc, %	4.6	4.6	9.2	9.2
Regenerant Flow Rate, gpm/cu ft	0.50	0.50	0.25	0.25
Volume Capacity, gals/cu ft	380	310	309	230
AMD Metal Ion Conc, gpg ($CaCO_3$)	69.2	67.1	68.4	64.3
Exchange Capacity, Kgrs/cu ft	26.3	20.8	21.2	14.8
% Regenerant Utilization	46	54	55	77
Lbs Waste Acid/Kilograin	0.12	0.09	0.09	0.03
Composite Treated Effluent				
Ferrous Iron, ppm Fe	3.7	5.5	5.0	20
Ferric Iron, ppm Fe	0.2	1.2	0.1	2
Calcium, ppm Ca	3.5	6.5	5.0	20
Magnesium, ppm Mg	0.5	0.9	0.7	2.9
Aluminum, ppm Al	0.0	0.0	0.0	0.0
Manganese, ppm Mn	0.1	0.2	0.1	0.7
Sodium, ppm Na	0.8	0.5	3.0	0.8
FMA, ppm $CaCO_3$	1500	1600	1500	1400

* Lbs of absolute (100%) hydrochloric acid per cubic foot of exchanger.

composite treated effluent. This is not unusual because the regenerant flow rate was halved when the concentration was doubled which resulted in the same regenerant contact time with the ion exchanger bed. The higher concentration would be preferred if hydrochloric acid regenerations were to be used because a smaller volume of waste effluent would be produced. The regeneration waste products produced by a hydrochloric acid regeneration are highly soluble chloride salts. Thus, the higher concentrations may be used with this regenerant.

The effect of acid dosage upon capacity follows the usual trend of ion exchange processes. Decreasing dosage causes a decrease in ion exchange capacity and an increase in regenerant utilization. The use of an hydrochloric acid dosage of 2 lbs/cu ft achieves a regenerant utilization of 77%.

Table 17 compares the results obtained in Section 8 using 3 lbs of sulfuric acid per cubic foot (Run 2A) with those obtained here using 2 lbs of hydrochloric acid per cubic foot. The characteristics of the composite effluent produced by these two acids are very similar. The

TABLE 17

Strong Acid Cation Exchanger - Hydrogen Form.
Performance Data.

Run No.	2A	77B	
Regenerant	H_2SO4	HCl	
Regenerant Dosage, lbs/cu ft	3	2	
Regenerant Dosage, Kgrs ($CaCO_3$)/cu ft	21.4	19.2	
Volume Capacity, gals/cu ft	153	230	
AMD Metal Ion Conc., gpg ($CaCO_3$)	70.0	64.3	
Exchange Capacity, Kgrs/cu ft	10.7	14.8	
% Regenerant Utilization	50	77	
Lbs Waste Acid/Kilograin	0.14	0.03	
Composite Treated Effluent			Typical Raw AMD
Ferrous Iron, ppm Fe	25.0	20.0	204
Ferric Iron, ppm Fe	2.5	2.0	13
Calcium, ppm Ca	15.0	20	182
Magnesium, ppm Mg	3.3	2.9	26
Aluminum, ppm Al	N.A.	0.0	18
Manganese, ppm Mn	1.0	0.7	8.5
Sodium, ppm Na	28.0	0.8	18
FMA, ppm $CaCO_3$	1430	1400	600

significant aspect of this comparison is the fact that the regenerant utilization is much higher when hydrochloric acid is used for regeneration. The greater regenerant utilization and higher exchange capacity obtained with hydrochloric acid also results in a much smaller quantity of unused acid discharged to waste per kilograin of exchange capacity than would occur if sulfuric acid regenerant were used.

The encouraging result obtained by the use of hydrochloric acid regenerant with AMD solutions containing essentially 100% ferrous iron led to the study of the same regenerant on AMD solutions containing 100% ferric iron. Table 18 compares the data obtained by the use of hydrochloric acid and sulfuric acid regenerants with AMD solutions containing nearly 100% ferric iron. The data in this table indicates that the exchange capacity obtained with this type of AMD solution is slightly higher than that obtained with AMD solutions containing 100% ferrous iron (Table 17). This result could be due to the greater selectivity of the cation exchanger for ferric iron over ferrous iron.

TABLE 18

Strong Acid Cation Exchanger - Hydrogen Form.
Performance Data.

Run Nos.	83B-86B	83A-86A	
Regenerant	H_2SO_4	HCl	
Regenerant Dosage, lbs/cu ft	3	2	
Regenerant Dosage, Kgrs ($CaCO_3$)	21.4	19.2	
Volume Capacity, gals/cu ft	185	230	
AMD Metal Ion Conc. gpg ($CaCO_3$)	74.2	74.2	
Ion Exchange Capacity, Kgrs/cu ft	13.5	17.0	
% Regenerant Utilization	63	89	
Lbs Waste Acid/Kilograin	0.08	0.013	
Composite Treated Effluent			Typical Raw AMD
Ferrous Iron, ppm Fe	0.6	0.7	0.3
Ferric Iron, ppm Fe	7.4	10.8	200
Calcium, ppm Ca	26	38	203
Magnesium, ppm Mg	6.6	6.9	27
Aluminum, ppm Al	0.0	0.0	17
Manganese, ppm Mn	1.5	1.6	8.8
Sodium, ppm Na	0.8	0.7	0.5
FMA, ppm $CaCO_3$	1560	1520	910

A comparison of the character of the composite treated effluents in Tables 17 and 18 is sufficient evidence of this. Effluents are lower in total iron but higher in calcium and magnesium when the AMD contains ferric iron rather than ferrous iron. This is the expected result because ferric iron removal is superior to ferrous iron removal. As a consequence. greater leakage of calcium and magnesium appears before unacceptable iron concentrations in the effluent.

Table 18 illustrates again the greater percentage of regenerant utilization which can be achieved by the use of hydrochloric acid regeneration. Only minor differences in effluent character are observed from the use of this regenerant also.

Because of the results obtained above. a comparison was made of the chemical costs (regenerant costs) which would be required to achieve 1000 grains (1 kilograin) of ion exchange capacity. These data are shown in Table 19 and were rather discouraging to the use of hydrochloric acid. The costs of using this acid would be 45-60% higher than that incurred by the use of sulfuric acid. This figure would represent

TABLE 19

Strong Acid Cation Exchanger - Hydrogen Form.
Performance Data.

AMD Iron	100% Ferrous		100% Ferric	
Regenerant	H_2SO_4	HCl	H_2SO_4	HCl
Regenerant Dosage, lbs/cu ft	3	2	3	2
Capacity, Kgrs/cu ft	10.7	14.8	13.5	17.0
% Regenerant Utilization	50	77	63	89
Waste Regenerant, lbs/Kgrs	0.14	0.03	0.08	0.013
Regenerant cost, ¢/Kgr*	0.46	0.67	0.37	0.59

*Based upon a sulfuric acid cost of $33.00/ton (1.65¢/lb) and a hydrochloric acid cost of $100/ton (5¢/lb) - 100% basis.

an increased cost of about 16¢ per thousand gallons when treating acid mine drainage having the same characteristics as the synthetic AMD used in this study. This cost increase would be considered unacceptable in view of the additional chemical costs which would be required to finish the treatment (weak base anion treatment).

An additional effort was made toward increasing the regenerant utilization of either the hydrochloric acid regeneration or the sulfuric acid regeneration. An attempt was made to recover waste regenerant from one cation exchanger column and reuse it in a preliminary regeneration process of a second column. The regeneration of the second column was then completed by the application of additional quantities of fresh regenerant.

Table 20 summarizes results of the regenerant recovery studies. In the case of hydrochloric acid regenerations, recovery procedures do increase the capacity of the ion exchanger column to which the recovered acid is applied. However, regenerant utilization is not increased when considering the total quantity of regenerant used for both columns and the total capacity obtained from both columns.

In the case of the sulfuric acid regeneration, the capacity of the column to which the waste regenerant was applied was actually lower than when only fresh regenerant contacted the column. This may have been caused by precipitation of calcium sulfate in the cation exchanger bed when in contact with the waste regenerant. The recovery procedure was carried out by passing the waste regenerant from the first column directly into the second column.

Calcium sulfate post-precipitation has been observed to occur in some of the sulfuric acid waste effluents. It is possible that this compound

TABLE 20

Strong Acid Cation Exchanger - Hydrogen Form.
Performance Data.

Run No.	77B	71A	69B	74B	86B	89B
Regenerant	HCl	HCl	HCl	HCl	H_2SO_4	H_2SO_4
AMD Iron	100% Ferrous				100% Ferric	
Regenerant Dosage						
Waste, lbs/cu ft	0	2	4	4	0	3
Fresh, lbs/cu ft	2	4	2	2	3	3
Regenerant Conc, %	9.2	4.6	4.6	9.2	2.0	2.0
Capacity, Kgrs/cu ft	14.8	21.1	17.4	17.8	13.5	12.5
% Regenerant Utilization	77	55	92	92	63	58
% Regenerant Utilization (overall)*	77	68	66	68	63	61

*Based upon the capacity and dosage used to regenerate the column from which the waste regenerant was obtained and the column to which the waste regenerant and fresh regenerant was applied.

could have post-precipitated from the waste regenerant as it was applied to the second column. This only points up the fact that the sulfuric acid waste regenerant has no further value from an ion exchanger regeneration standpoint.

Tables 82 through 87 show the character of the effluent during each of the treatment cycles cited here as typical of each variable set. These data show the variation of the effluent cation concentration which occurs during the treatment cycle.

The results of the study of the strong acid cation exchanger (hydrogen form) have indicated that hydrochloric acid regeneration is impractical from an economic standpoint if only regenerant cost is considered. However, a consideration of the ultimate disposal of waste regenerants could alter this opinion. Regenerant waste disposal is not within the scope of this study and some method of disposal will be required if this AMD treatment process is to become a reality. If the regenerant waste is to be neutralized and/or hauled away, the neutralizing chemical cost and hauling costs will be an important economic consideration. The fact that hydrochloric acid regeneration results in substantially less waste acidity and waste regenerant volume may bring this material back into the picture.

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SECTION 14

COMPLETE PROCESS EVALUATION - STRONG ACID CATION EXCHANGER (HYDROGEN FORM)/WEAK BASE ANION EXCHANGER (FREE BASE FORM)

This complete process utilizes the combination of two separate ion exchange processes. The first step of the complete process incorporates the hydrogen form of a strong acid cation exchanger. A study of this process was carried out as described in Sections 8 and 13. The effluent from the cation exchanger is treated further by the use of a weak base anion exchanger in the free base form and this process is described in Section 10 of this report.

It was recognized during the studies of the weak base anion exchanger that the effluent from the weak base anion exchanger would not meet the requirements for potability. This failure would be due to the presence of excessive concentrations of iron and manganese which are abundantly present in the cation exchanger effluent and are not completely removed by the weak base anion exchanger. One of the objectives of this study was to establish a post treatment method which would render the water fit for human consumption. Aeration at elevated pH is a known procedure for removal of iron and manganese. The complete treatment process utilizing a strong acid cation exchanger followed by a weak base anion exchanger followed by pH elevation, aeration and filtration was evaluated in this study. A schematic representation of this process is shown in Figure 4.

This evaluation involved the application of the complete process to the treatment of acid mine drainage containing either 100% ferrous iron or 100% ferric iron. During this study, raw AMD solutions were used to backwash the cation exchanger, to prepare the sulfuric acid regenerant solution and to rinse the cation exchanger after applying the regenerant. The cation exchanger effluent was used to backwash the weak base anion exchanger, prepare the sodium hydroxide regenerant and rinse the weak base anion exchanger after applying the regenerant.

U. S. Department of Health, Education and Welfare potability specifications require that the manganese content be no greater than 0.05 ppm and the iron content be no greater than 0.3 ppm. Because oxidation and precipitation of manganese is pH dependent, an investigation was conducted to determine the minimum pH required for adequate manganese and iron removal. Table 21 shows results of this investigation and indicates that a minimum pH of 9.9 is needed to remove manganese to the desired level. Data for Table 21 was obtained from water produced by the addition of lime to the weak base anion effluent until the desired pH was attained. Then, the solution was aerated, the precipitate allowed to settle and the supernatant filtered through a 5 micron filter. The iron and manganese analysis was performed on the filtered sample.

TABLE 21

Effect of pH on Manganese and Iron Removal Using
Lime Treatment and Filtration

pH Levels	7.60	9.50	9.95	9.90
Ferrous Iron, ppm Fe	0.0	0.0	0.0	0.0
Ferric Iron, ppm Fe	0.1	0.0	0.0	0.0
Manganese, ppm Mn	0.17	0.17	0.5	0.04

Table 22 shows the chemical characteristics of the raw AMD and the various effluents of each stage of the complete treatment process when the influent contained essentially 100% ferrous iron. Table 23 is an example of the various effluent characteristics obtained by the treatment of AMD containing essentially 100% ferric iron. The additional pH correction step was added (as shown in Table 23) because the high pH resulting from the manganese removal process would not be acceptable from a potability standpoint either. The correction of pH was accomplished here by blending a portion of the cation exchanger effluent with the filter effluent. The ratio of this blend was one volume of cation effluent to 100 volumes of filter effluent. This method of pH correction is shown to be acceptable because it produces negligible increases in iron and manganese.

Tables 88 through 90 show the capacity and final product quality obtained by each of the five complete treatment cycles produced with AMD containing 100% ferrous iron and by each of the fourteen complete treatment cycles produced with AMD containing 100% ferric iron. The analyses were made on the final treated product of the complete process. However, none of the data shows the final product after pH correction to the neutral range. It has been demonstrated before that the final pH correction process would not materially alter the chemical characteristics of the water.

It should be noted that repetitive cycling produced an average of 36.0 Kgrs/cubic foot capacity for the weak base anion exchanger (overall average of the last eighteen runs omitting the first run). A capacity of 30.0 Kgrs/cubic foot was selected as the plant design criteria allowing for some capacity losses to occur before having to institute resin clean-up or replacement procedures. There was no indication of a capacity decline over the nineteen cycles produced during this study. However, this would be the expected result of exhausting a weak base anion exchanger with a hydrogen form strong acid cation exchanger effluent rather than a raw AMD solution. The total iron content of the weak base anion exchanger influent (hydrogen form cation exchanger effluent) was typically 20-30 ppm with a 100% ferrous iron cation exchanger influent and was 3-6 ppm with a 100% ferric iron cation exchanger influent.

TABLE 22 (Run No. 98)

Typical Effluent Analysis During Each Step of the AMD Treatment (100% Ferrous) Using a Strong Acid Cation Exchanger (Hydrogen Form), a Weak Base Anion Exchanger (Free Base Form), Liming and Filtration.

	Raw AMD	Cation Effluent	Anion Effluent	After liming, settling & filtration
pH	2.35	1.85	8.0	9.90
Iron, ppm Fe^{++}	190	10.2	0.0	0.0
Iron, ppm Fe^{+++}	10	1.8	0.9	0.0
Magnesium, ppm Mg^{++}	27	1.5	1.5	1.5
Calcium, ppm Ca^{++}	205	13	13	17
Manganese, ppm Mn^{++}	8.9	0.47	0.34	0.04
Aluminum, ppm Al^{+++}	15	0	0	0
Sodium, ppm Na^{++}	0.5	8.7	23	23
Sulfate, ppm $SO_4^{=}$	1850	1611	62	61
Alkalinity, ppm $CaCO_3$	--	--	25	35
FMA, ppm $CaCO_3$	945	1540	--	--

TABLE 23 (Run No. 105)

Typical Effluent Analysis During Each Step of the AMD Treatment (100% Ferric) Using a Strong Acid Cation Exchanger (Hydrogen Form), A Weak Base Anion Exchanger (Free Base Form) Liming, Filtration and pH Correction.

	Raw AMD	Cation Effluent	Anion Effluent	After liming, settling & filtration	After pH Correction
pH	2.10	1.65	9.5	10.1	7.5
Iron, ppm Fe^{++}	0.3	0.3	0.1	0.0	0.0
Iron, ppm Fe^{+++}	210	2.9	0.5	0.0	0.1
Magnesium, ppm Mg^{++}	28	2.4	2.2	2.1	2.1
Calcium, ppm Ca^{++}	200	20	28	18	18
Manganese, ppm Mn^{++}	8.9	0.77	0.18	0.00	0.01
Aluminum, ppm Al^{++}	14	0	0	0	0
Sodium, ppm Na^{+}	0.9	2.9	39	39	38
Sulfate, ppm $SO_4^{=}$	1990	2000	129	95	113
Alkalinity, ppm $CaCO_3$	--	--	30	40	20
FMA, ppm $CaCO_3$	1440	2020	--	--	--

Tables 91 through 93 show the analysis of the hydrogen form strong acid cation exchanger effluents. No cation exchanger capacity data was acculated during this study since this information has been established by the study described in Section 13.

Studies should be conducted to investigate the effect of the iron accumulation upon the weak base anion capacity over several hundred cycles. The usable resin life would thereby be established for AMD treatment. In turn, the weak base anion amortization rate would be a factor in the operating costs for a full scale plant.

During the course of this evaluation the information shown in Table 24 was obtained. The data in Table 24 reflect the results of studies carried out to determine the minimum rinse requirement for the cation and anion exchanger columns. It was found that 7.5 gallons per cu ft of anion exchanger was adequate to displace the regenerant from the interstices of the ion exchanger bed and also to reduce any residual regenerant to an acceptable level. It should be noted that this volume of rinse would be inadequate for most demineralization processes where waters with substantially lower dissolved solids are to be produced. However, in this case, the process is attempting only to produce a potable water which may contain relatively high concentrations of dissolved solids. Thus, it should be permissible to use the minimum quantity of rinse necessary to reduce the dissolved solids only to a potable level. Such a procedure would minimize the volume of waste effluent and reduce the ionic load to the ion exchanger. An additional benefit is obtained by allowing some of the sodium hydroxide regenerant from the anion exchanger to appear in the treated water because this effluent will eventually require pH elevation for manganese removal. Thus the quantity of chemical needed to raise the pH would be diminished.

On the basis of the capacity data obtained in this study for the strong acid cation exchanger and weak base anion exchanger, the chemical costs required to obtain 1,000 gallons of treated effluent from AMD having the chemical characteristics shown in Table 22 are

Chemical	Lbs/1000 Gals	@	¢/lb	=	¢/1000 Gals
Sulfuric Acid	17.1		1.57		26.9
Caustic Soda	9.8		3.7		36.2
Lime	0.1		1.0		0.1
			Total		63.2¢

Sulfuric acid costs are based upon a tank car cost of \$31.50 per ton and caustic soda costs are based upon a tank car 50% liquid cost of \$3.70 per 100 lbs (78% Na_2O-100% NaOH basis).

TABLE 24

Ion Exchange Column Operational Parameters - Strong Acid Cation Exchanger (Hydrogen Form) - Weak Base Anion Exchanger (Free Base Form) Treatment Process

PROCESS	Cation Exchanger	Anion Exchanger
Ion Exchanger Type	Strong Acid	Weak Base
Regenerant	Sulfuric Acid	Caustic Soda
Concentration	2%	4%
Dosage, lbs/cu ft	3.0	3.5
Flow Rate, gpm/cu ft	0.5	0.5
Direction	Downflow	Downflow
Rinse	AMD	Cation Effluent
Flow Rate, gpm/cu ft	0.5	0.5
Volume, gals/cu ft	7.5	7.5
Backwash	AMD	Cation Effluent
Flow Rate, gpm/cu ft	Depends upon unit dimension.	
Volume, gals/cu ft	Depends upon character of AMD.	
Treatment		
Flow Rate, gpm/cu ft	4.0	2.0
Direction	Downflow	Downflow
Capacity, Kgrs/cu ft	12.5	30.0

Figure 8 represents the material balance for the complete treatment process with an input of 100,000 gallons of raw AMD. This illustration does not attempt to show the exact quantities of waste materials which result from water producing reactions. For example, the regeneration of the weak base anion exchanger involves the following reaction:

$$\underline{R\text{-}NH_2 \cdot H_2SO_4} + 2NaOH \longrightarrow \underline{R\text{-}NH_2} + Na_2SO_4 + 2H_2O. \qquad (1)$$

The water produced by this reaction is included with the solid waste weight in Figure 8. Thus, the actual dehydrated weights of waste material may be somewhat lower than that shown.

Although the ultimate disposal of the waste materials is not within the scope of this study, the fate of these materials is vital to the success of this process. It would be difficult to speculate on a method for treatment and/or disposal since it seems likely that such methods would be governed by local waste disposal laws in the area to which the process is to be applied.

For the most part, the solid wastes are water soluble. The acid waste contains excess acids which if neutralized by lime, would produce some insoluble wastes. It seems likely that the remaining water soluble wastes (in solution) would either have to be hauled to an approved disposal site (if available) or dewatered and the solid material disposed of in some manner (not known).

The economic aspects of neutralizing and/or hauling the relatively large volume of acid wastes from the sulfuric acid regeneration could wipe out its cost advantage over the hydrochloric acid regeneration. The regeneration with hydrochloric acid would produce approximately 5000 gallons of acid wastes vs the 12,850 gallons produced by the sulfuric acid regeneration (based upon 100,000 gallons AMD input). Moreover, the amount of excess acidity in the waste would be less with hydrochloric acid regeneration because it is more efficient (reference Section 13). Hydrochloric acid wastes may be neutralized with calcium carbonate because no insoluble materials are formed by this reaction. It may be possible to neutralize the hydrochloric acid wastes partially with calcium carbonate and then finish the neutralizing process using the residual caustic present in the caustic waste from the weak base anion exchanger regeneration. Thus a neutral waste effluent containing mixed calcium, magnesium, manganese, sodium, sulfate and chloride salts could be achieved. We would expect iron to precipitate as the hydroxide and some calcium sulfate to precipitate when the sodium sulfate wastes (from the anion regeneration) are combined with the calcium chloride wastes (from the cation regeneration).

It is recommended that the economics of the hydrochloric acid and sulfuric acid regeneration be compared through the application of a pilot plant on an actual AMD source. The comparison should consider the requirement for waste treatment and disposal.

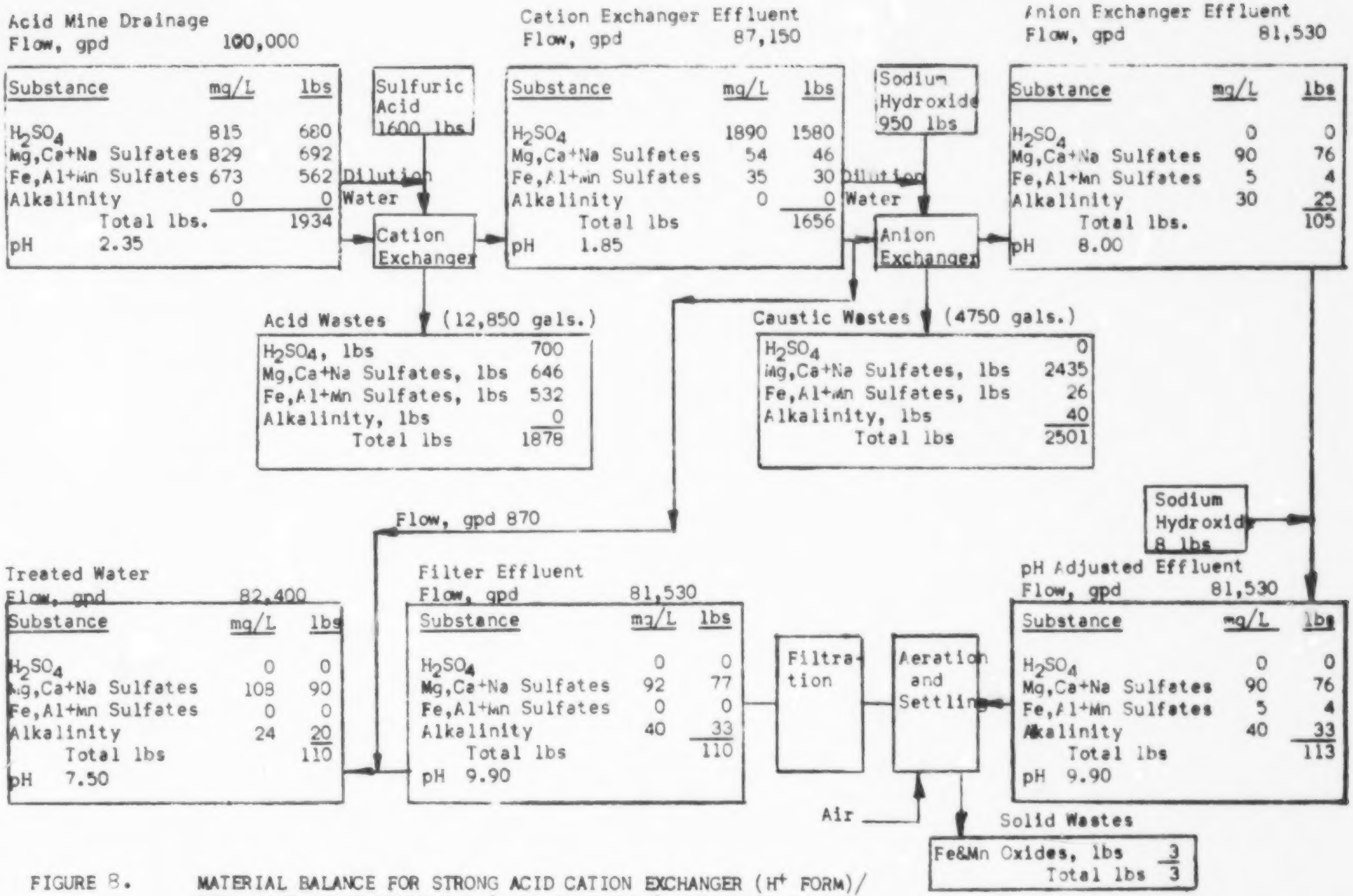
Acid Mine Drainage
Flow, gpd 100,000
Substance mg/L lbs
H2SO4 815 680
Mg,Ca+Na Sulfates 829 692
Fe,Al+Mn Sulfates 673 562
Alkalinity 0 0
Total lbs. 1934
pH 2.35
Sulfuric Acid 1600 lbs
Dilution Water
Cation Exchanger
Acid Wastes (12,850 gals.)
H2SO4, lbs 700
Mg,Ca+Na Sulfates, lbs 646
Fe,Al+Mn Sulfates, lbs 532
Alkalinity, lbs 0
Total lbs 1878
Cation Exchanger Effluent
Flow, gpd 87,150
Substance mg/L lbs
H2SO4 1890 1580
Mg,Ca+Na Sulfates 54 46
Fe,Al+Mn Sulfates 35 30
Alkalinity 0 0
Total lbs 1656
pH 1.85
Sodium Hydroxide 950 lbs
Dilution Water
Anion Exchanger
Caustic Wastes (4750 gals.)
H2SO4 0
Mg,Ca+Na Sulfates, lbs 2435
Fe,Al+Mn Sulfates, lbs 26
Alkalinity, lbs 40
Total lbs 2501
Anion Exchanger Effluent
Flow, gpd 81,530
Substance mg/L lbs
H2SO4 0 0
Mg,Ca+Na Sulfates 90 76
Fe,Al+Mn Sulfates 5 4
Alkalinity 30 25
Total lbs. 105
pH 8.00
Sodium Hydroxide 8 lbs
pH Adjusted Effluent
Flow, gpd 81,530
Substance mg/L lbs
H2SO4 0 0
Mg,Ca+Na Sulfates 90 76
Fe,Al+Mn Sulfates 5 4
Alkalinity 40 33
Total lbs 113
pH 9.90
Aeration and Settling
Air
Solid Wastes
Fe&Mn Oxides, lbs 3
Total lbs 3
Filtration
Filter Effluent
Flow, gpd 81,530
Substance mg/L lbs
H2SO4 0 0
Mg,Ca+Na Sulfates 92 77
Fe,Al+Mn Sulfates 0 0
Alkalinity 40 33
Total lbs 110
pH 9.90
Flow, gpd 870
Treated Water
Flow, gpd 82,400
Substance mg/L lbs
H2SO4 0 0
Mg,Ca+Na Sulfates 108 90
Fe,Al+Mn Sulfates 0 0
Alkalinity 24 20
Total lbs 110
pH 7.50

FIGURE 8. MATERIAL BALANCE FOR STRONG ACID CATION EXCHANGER (H^+ FORM)/ WEAK BASE ANION EXCHANGER (FREE BASE FORM) TREATMENT OF ACID MINE DRAINAGE

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SECTION 15

COMPLETE PROCESS EVALUATION
WEAK BASE ANION EXCHANGER (BICARBONATE FORM)/LIME TREATMENT

This complete treatment process utilizes the special weak base anion exchanger (Rohm & Haas' IRA-68) in the bicarbonate form. This process was discussed briefly in Section 11 and was found to produce an effluent containing essentially bicarbonate salts of the divalent metal ions which were originally present in the raw AMD. A complete treatment process for the production of a potable water would require additional treatment of the weak base anion effluent.

Two alternative treatment methods are available for completing this process. One treatment involves the addition of lime to precipitate iron, calcium, magnesium and manganese (equations 3, 4, 5 and 6 of Section 11). A second treatment method utilizes a weak acid cation exchanger in the hydrogen form to absorb only divalent metal ions associated with alkalinity. Weak acid cation exchangers do not have the ability to "split" neutral salts. Thus, these compounds remain unchanged in solution after being subjected to this treatment. Equation 1 illustrates the reaction which occurs with alkaline (bicarbonate) salts.

$$Ca(HCO_3)_2 + \underline{2R\text{-}COOH} \rightarrow \underline{(R\text{-}COO)_2Ca} + 2H_2CO_3 \qquad (1)$$

The carbonic acid formed in equation 1 can dissociate to form water and carbon dioxide in subsequent aeration processes (equation 2).

$$H_2CO_3 \rightarrow H_2O + CO_2 \qquad (2)$$

The regeneration process for the weak acid cation exchanger is carried out by contacting the exchanger with a strong mineral acid solution such as sulfuric acid (equation 3).

$$\underline{(R\text{-}COO)_2Ca} + H_2SO_4 \rightarrow \underline{2R\text{-}COOH} + CaSO_4 \qquad (3)$$

As was necessary when regenerating the strong acid cation exchanger, the sulfuric acid regenerant solution must be maintained at 2% or less in order to prevent calcium sulfate precipitation. Because of the favorable selectivity for hydrogen ion by the weak acid cation exchanger, regenerant utilization is very nearly 100%. It is customary to apply 110% of the theoretical acid dosage to weak acid cation exchangers.

Figure 7 is a schematic representation of these two complete processes. Post treatment No. 1 consists of the weak acid cation treatment and subsequent lime treatment of the IRA-68 effluent. Post treatment No. 2 consists of aeration, liming and filtration of the IRA-68 effluent.

TABLE 25

Typical Effluent Analyses During Each Step of the AMD Treatment with a Weak Base Anion Exchanger (Bicarbonate Form), a Weak Acid Cation Exchanger (Hydrogen Form), Aeration, Liming and Filtration (Run 121)

	Untreated AMD	Weak Base Anion Effluent	Weak Acid Cation Effluent	After aeration, liming & filtering
pH	2.45	6.80	4.05	10.9
Ferrous Iron, ppm Fe	210	90	0.4	0.0
Ferric Iron, ppm Fe	10	20	0.2	0.0
Calcium, ppm Ca	190	185	0.7	17
Magnesium, ppm Mg	28	28	0.4	0.3
Aluminum, ppm Al	15	5	1	1
Manganese, ppm Mn	8.3	8.1	0.06	0.03
Sodium, ppm Na	0.9	1.8	1.2	1.1
Sulfate, ppm SO_4	1430	105	13	6
Alkalinity, ppm $CaCO_3$	--	500	--	40
FMA, ppm $CaCO_3$	490	--	5	--
Ion Exchange Capacity Kgrs/cu ft		25.2*	7.7**	
% Regenerant Utilization		91	26	

*Based upon a loading factor of 58.0 grains/gallon (influent FMA concentration plus effluent alkalinity concentration).

**Based upon a loading factor of 29.0 grains/gallon (influent alkalinity concentration).

TABLE 26

Weak Acid Cation Exchanger Capacity and Regenerant Utilization vs Sulfuric Acid Dosage

Sulfuric Acid Dosage, lbs/cu ft	2.05	4.1	8.2
Sulfuric Acid Dosage, Kgrs/cu ft	14.6	29.2	58.4
Ion Exchange Capacity, Kgrs/cu ft	7.1	7.2	7.4
% Regenerant Utilization	49	25	13

The weak acid cation exchanger used in the No. 1 post treatment study was manufactured by the Rohm & Haas Company and is designated by them as IRC-84. The Rohm and Haas Company recommends that regeneration of this exchanger be accomplished by passing a 0.5% solution of sulfuric acid downflow through the ion exchanger bed at a flow rate of 1 gpm/cu ft. A 30 minute rinse was applied at the same flow rate and direction after the regenerant was passed through.

The IRA-68 ion exchanger was regenerated with sodium hydroxide and converted to the bicarbonate form using the same procedure as was used in the study described in Section 11.

When a lime treatment was applied, the lime was added either at the beginning or just prior to the end of the aeration process. Then, the mixture was allowed to settle for a 30 minute period followed by passage of the supernatant through a 5 micron filter.

Treatment processes through the ion exchanger columns were carried out at flow rates of 2 gpm/cu ft. The weak base anion exchanger was operated upflow and the weak acid cation exchanger (when used) was operated downflow during the treatment cycle.

This evaluation has demonstrated that the process involving the weak acid cation exchanger treatment has the ability to produce a potable effluent. However, this process is more costly than the treatment involving only aeration, liming and filtration. This may have been due in part to our inability to achieve the regenerant efficiency which is predicted for the weak acid cation exchanger.

The process involving aeration and lime treatment also has the ability to produce a potable product and has shown to be comparable to the strong acid cation/weak base anion process insofar as chemical costs are concerned.

Table 25 shows the characteristics of the waters before and after each treatment step when post treatment No. 1 (weak acid cation exchanger) was used. These data illustrate the ability of this treatment process to produce a potable effluent provided a final pH correction step was applied. Although the process appears successful from the water quality standpoint, the capacity obtained by the weak acid cation exchanger was approximately 1/4 that which would be expected from the amount of regenerant applied. This ion exchanger has an operating capacity of about 60 Kgrs/cu ft. Thus, there was no concern that we were operating too near its maximum capacity.

A study was made to determine the effect on capacity by variations of the sulfuric acid regenerant dosage. Table 26 shows results of this study and indicates that variations of regenerant dosage have no affect on capacity. These data indicate that the weak acid cation exchanger probably does not have the kinetic capabilities to handle the alkalinity load imposed by this type of water at practical flow

TABLE 27 (Run 122)

Typical Effluent Analyses During Each Step of the AMD Treatment With A Weak Base Anion Exchanger(Bicarbonate Form), Aeration, Liming and Filtration (100% Ferrous)

	Raw AMD	Weak Base Effluent	After Aeration, Liming Settling & Filtering
pH	2.5	6.8	10.1
Iron, ppm Fe^{++}	179	98	0.0
Iron, ppm Fe^{+++}	21	32	0.0
Magnesium, ppm Mg^{++}	28	27	15
Calcium, ppm Ca^{++}	180	180	17
Manganese, ppm Mn^{++}	8.0	7.8	0.01
Aluminum, ppm Al^{+++}	15	4	0.0
Sodium, ppm Na^{+}	0.9	1.3	1.3
Sulfate, ppm $SO_4^{=}$	1460	75	70
Alkalinity, ppm $CaCO_3$	--	675	35.0
FMA, ppm $CaCO_3$	495	--	--

TABLE 28 (Run 128)

Typical Effluent Analysis During Each Step of the AMD Treatment With A Weak Base Anion Exchanger (Bicarbonate Form), Aeration, Liming and Filtration (100% Ferric)

	Raw AMD	Weak Base Efflunet	After Aeration, Liming Settling & Filtering
pH	2.3	6.3	9.9
Iron, ppm Fe^{++}	0.5	0.1	0.0
Iron, ppm Fe^{+++}	200	48	0.0
Magnesium, ppm Mg^{++}	28	27	9
Calcium, ppm Ca^{++}	200	200	16.0
Manganese, ppm Mn^{++}	7.7	6.8	0.00
Aluminum, ppm Al^{+++}	13	4	0
Sodium, ppm Na^{+}	1.0	2.0	2.0
Sulfate, ppm $SO_4^{=}$	1670	93	54
Alkalinity, ppm $CaCO_3$	--	540	25.0
FMA, ppm $CaCO_3$	1105	--	--

rates. In view of these results, the study of this post treatment process was abandoned.

The treatment process involving aeration, liming and filtration after weak base anion exchange (bicarbonate form) is capable of producing a potable effluent as presented in Tables 27 and 28. Table 27 shows the character of the effluents before and after each treatment step when the influent (raw AMD) iron was essentially 100% ferrous, while Table 28 shows the results obtained when the influent iron was nearly 100% ferric. Product water in either case was suitable for human consumption with the exception of pH. Correction of pH could be accomplished simply by feeding approximately 10-15 ppm of an acid.

Tables 94 through 97 show the characteristics of the final treated effluent (without pH correction) produced by successive runs on each complete treatment process. Table 94 shows the results of bicarbonate form weak base anion - lime treatment with an AMD influent containing essentially 100% ferrous iron. The treated effluent obtained on each of the six cycles would be suitable for human consumption after pH correction. A capacity in excess of 30.0 Kgrs/cu ft was obtained.

Table 95 shows results of the same treatment process when treating an AMD containing essentially 100% ferric iron. Potable treated effluents were obtained from each of the six runs also. However, a lower capacity of the weak base anion exchanger was observed. Subsequent cost calculations and plant designs were based upon the 30.0 Kgr/cu ft capacity value. However, application of this process to waters containing large concentrations of ferric iron could require a slight reassessment of the size of the weak base anion exchanger quantities and the cost of producing the treated water.

Tables 96 and 97 show results of the complete treatment process utilizing a bicarbonate form weak base anion exchanger followed by a hydrogen form weak acid cation exchanger on AMD solutions containing 100% ferrous iron and 100% ferric iron respectively. In all runs the weak acid cation effluent was limed to a pH of 9.9 or above to precipitate manganese. No pH correction was carried out. Since this treatment process was not an economic success, this data is presented for academic interest only.

It was determined during the course of this investigation that considerable savings in lime consumption could be achieved by aerating the weak base anion effluent prior to the addition of lime. The reason for this is probably related to iron oxidation reactions which produce water and carbon dioxide (equation 4).

$$2Fe(HCO_3)_2 + \tfrac{1}{2}O_2 + 5H_2O \longrightarrow \underline{2Fe(OH)}_3 + 4H_2O + 4CO_2 \qquad (4)$$

If lime were added prior to aeration, some of the lime would react with the free CO_2 as illustrated by equation 5.

$$Ca(OH)_2 + CO_2 \rightarrow CaCO_3 + H_2O \qquad (5)$$

Although we were able to reduce the lime requirement by aerating prior to lime addition, we were not able to approach the theoretical requirement for lime. Our results have indicated that 15 pounds of lime would be needed to produce 1000 gallons of treated effluent whereas only 3.2 pounds should be needed at 100% reactivity. An investigation of this lime inefficiency was beyond the scope of the project.

A practical lime consumption estimate of 3.5 pounds per 1000 gallons of treated effluent has been calculated based upon calcium bicarbonate and magnesium bicarbonate concentrations only. It has been assumed that iron and aluminum will already be in the precipitated form following the aerator. Alkalinity will also have been diminished by an equivalent amount while free carbon dioxide content should be negligible.

Therefore, precipitation of calcium and magnesium from their bicarbonate salts requires respectively one and two equivalents of lime as shown in equations 6, 7 and 8.

$$CaO + H_2O \rightarrow Ca(OH)_2 \qquad (6)$$

$$Ca(HCO_3)_2 + Ca(OH)_2 \rightarrow \underline{2CaCO_3} + 2H_2O \qquad (7)$$

$$Mg(HCO_3)_2 + 2Ca(OH)_2 \rightarrow \underline{Mg(OH)_2} + \underline{2CaCO_3} + 2H_2O \qquad (8)$$

The requirements for lime (90% CaO) are calculated as follows, where ppm* is expressed as $CaCO_3$ equivalents.

$$\text{ppm* } Ca(HCO_3)_2 \times \frac{28 \text{ (eq wt CaO)}}{50 \text{ (eq wt } CaCO_3)} \times \frac{1}{0.9 \text{ (purity)}} = \text{ppm of 90\% CaO}$$

A dosage of one pound in 1000 gallons will provide a concentration of 120 ppm. Therefore, dividing ppm by 120 expresses the requirement in "pounds per 1000 gallons". Thus, each ppm of calcium (expressed as $CaCO_3$), will require 0.00519 pounds of 90% CaO per 1000 gallons.

$$1 \text{ ppm } Ca(HCO_3)_2 \times \frac{28}{50} \times \frac{1}{0.9} \times \frac{1}{120} = 0.0519$$

Each ppm of magnesium (expressed as $CaCO_3$) will require twice as much: or, 0.01038 pounds of 90% CaO per 1000 gallons. These factors for determining the lime requirements have been used in subsequent calculations involving this process.

Calcium concentration = 450 ppm $CaCO_3$

450 x 0.00519 = 2.34 lbs 90% CaO

Magnesium Concentration = 112 ppm $CaCO_3$

$112 \times 0.01038 = 1.16$ lbs 90% CaO

Total lime requirement per 1000 gal = 3.50 lbs 90% CaO

During the course of this investigation, the information displayed in Table 29 was obtained. The capacity of the IRA-68 was an average obtained over ten regeneration-exhaustion cycles treating an AMD solution containing essentially 100% ferrous iron and also containing 500-550 ppm of FMA. When synthetic AMD solutions containing 100% ferric iron were used, the FMA content was increased to 1000-1100 ppm in order to effect the solution of ferric sulfate. The capacity of the IRA-68 exchanger over 12 complete exhaustion-regeneration cycles with this AMD solution was only 23.3 Kgrs/cu ft. These values indicate that there will be some variations in capacity depending upon FMA loading. However, the 30 Kgr/cu ft value was selected for application to the standard AMD solution prepared with 100% ferrous iron.

On the basis of the capacity data obtained above, the approximate chemical costs required to obtain 1000 gallons of treated effluent from AMD having the chemical characteristics shown in Table 25 are as follows:

Chemical	Lbs/1000 Gallons	@ ¢/Lb.	¢/1000 Gals.
Caustic Soda	9.0	3.7	33.3
Carbon Dioxide	7.6	1.5	11.4
Lime	3.5	1.0	3.5
			48.2

Figure 9 represents the material balance for the complete treatment process with an input of 100,000 gallons. Of particular interest in this process is the amount and character of the waste products. The waste products from the anion exchanger are primarily soluble sulfate salts with some excess alkalinity. The waste products from the liming operation are insoluble sludge and water. No acidic wastes are produced by this process as was the case with the strong acid cation/weak base anion process. The solid wastes disposal would be relatively simple consisting probably of a dewatering step and disposal of the solids at an approved site. The disposal of the sodium sulfate solution is more difficult and a discussion of the problem may be found in Section 14.

TABLE 29

Ion Exchange Column Operational Parameters - Weak Base Anion Exchanger (Bicarbonate Form)/Lime Treatment

Process	Weak Base Anion Exchanger
Ion Exchange Type	IRA-68
Regenerants	
NaOH Dosage, lbs/cu ft	4.0
Concentration, %	4
Flow Rate, gpm/cu ft	0.5
Direction	Downflow
CO_2 Dosage, lbs/cu ft	3.3
Concentration	Variable
Flow Rate, gpm/cu ft	0.5
Direction	Downflow
Rinse	
Flow Rate, gpm/cu ft	0.5
Volume, gals/cu ft	7.5
Backwash	
Flow Rate, gpm/cu ft	*
Volume, gals/cu ft	**
Treatment	
Flow Rate, gpm/cu ft	2.0
Direction	Upflow
Capacity, Kgrs/cu ft	30.0
Lime	
Lbs/1000 gals	3.5

*Depends upon tank dimension.
**Depends upon influent AMD characteristics.

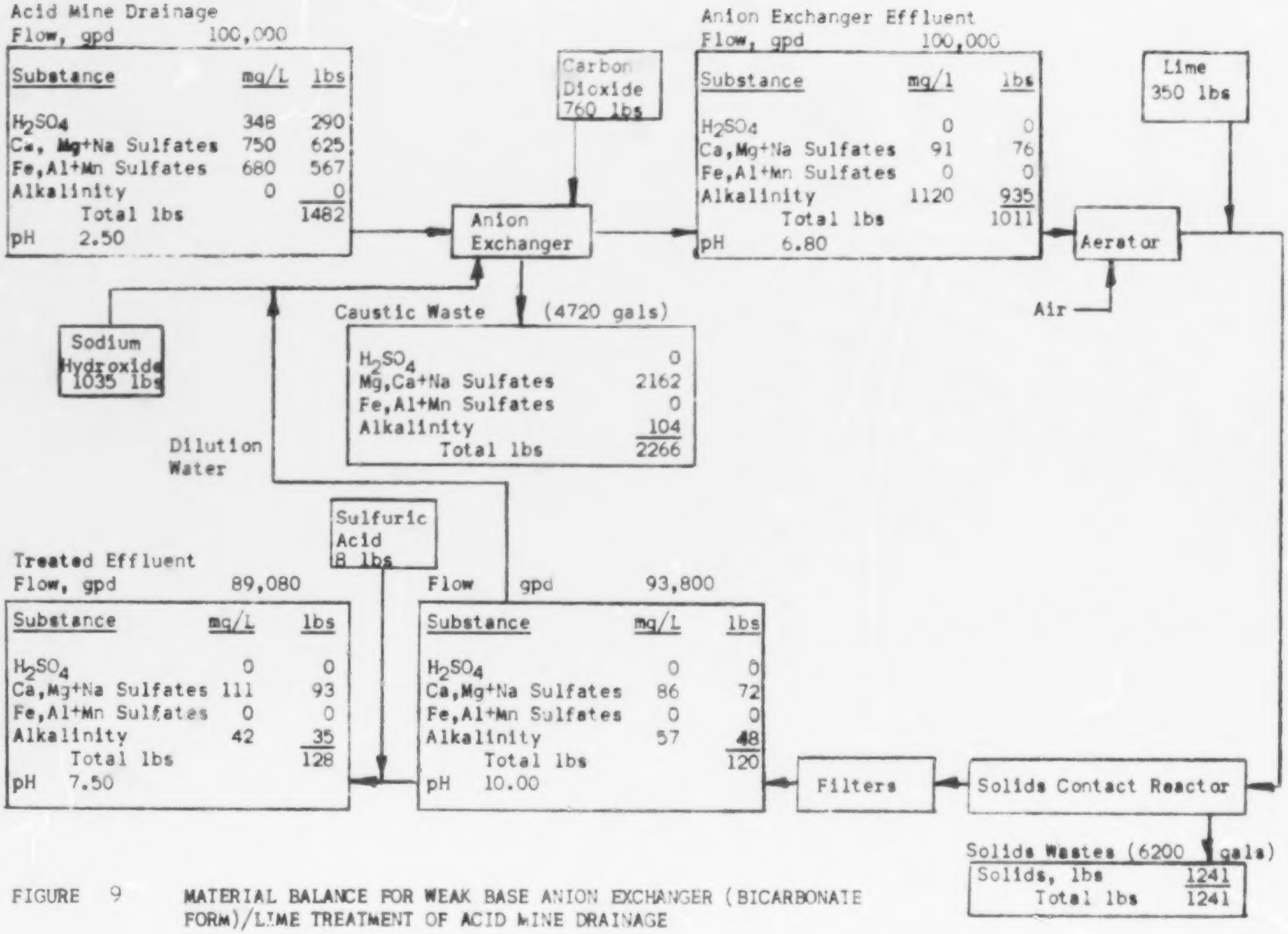
Acid Mine Drainage
Flow, gpd 100,000
Substance mg/L lbs
H_2SO_4 348 290
Ca, Mg+Na Sulfates 750 625
Fe,Al+Mn Sulfates 680 567
Alkalinity 0 0
Total lbs 1482
pH 2.50
Carbon Dioxide 760 lbs
Anion Exchanger
Sodium Hydroxide 1035 lbs
Dilution Water
Caustic Waste (4720 gals)
H_2SO_4 0
Mg,Ca+Na Sulfates 2162
Fe,Al+Mn Sulfates 0
Alkalinity 104
Total lbs 2266
Anion Exchanger Effluent
Flow, gpd 100,000
Substance mg/l lbs
H_2SO_4 0 0
Ca,Mg+Na Sulfates 91 76
Fe,Al+Mn Sulfates 0 0
Alkalinity 1120 935
Total lbs 1011
pH 6.80
Lime 350 lbs
Aerator
Air
Solids Contact Reactor
Solids Wastes (6200 gals)
Solids, lbs 1241
Total lbs 1241
Filters
Flow gpd 93,800
Substance mg/L lbs
H_2SO_4 0 0
Ca,Mg+Na Sulfates 86 72
Fe,Al+Mn Sulfates 0 0
Alkalinity 57 48
Total lbs 120
pH 10.00
Sulfuric Acid 8 lbs
Treated Effluent
Flow, gpd 89,080
Substance mg/L lbs
H_2SO_4 0 0
Ca,Mg+Na Sulfates 111 93
Fe,Al+Mn Sulfates 0 0
Alkalinity 42 35
Total lbs 128
pH 7.50

FIGURE 9 MATERIAL BALANCE FOR WEAK BASE ANION EXCHANGER (BICARBONATE FORM)/LIME TREATMENT OF ACID MINE DRAINAGE

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SECTION 16

TREATMENT PLANT DESIGN - TWO RESIN SYSTEM

The system considered in this section is designed to treat acid mine drainage to produce a water which will meet the mineral requirements for potable water. The system used two resins, employing a strongly acidic cation exchange resin followed by a weakly basic anion exchange resin and post treatment. Supporting laboratory work for this process has been previously discussed. Sections 8, 10 and 13 discussed the individual resins, while Section 14 discussed the complete process for production of potable water.

The design parameters for plants employing this process were presented in Table 24. Those parameters have been used to design plants for the treatment of acid mine drainage water. Plants of three sizes have been designed; namely, 0.1, 0.5, 1.0 MGD (million gallons per day).

Summary of Costs:

Cost estimates have been made for the three plants which were designed under this project. The costs are presented for each plant later in this section, but the costs are summarized here.

Figure 10 shows the effect of plant size on treatment costs for this two resin system as well as for the modified Desal system. The cost data were totaled from the individual costs detailed in this and the following section. Equipment, installation labor and ion exchange materials were amortized over ten years.

Figure 11 is a plot of the equipment costs, in hundreds of thousands of dollars for the various size plants to treat AMD by this two resin system. These prices are for the equipment listed in the detailed specifications, but exclude freight, building and land, assembly and erection.

Figure 12 is a plot of the estimated chemical operating costs, in hundreds of dollars per day for the plants to treat the AMD by the two resin system. All utility costs are excluded.

Figures 13 and 14 plot, respectively, the estimated erection labor costs for the electrical and plumbing requirements. These costs are based upon our plant designs.

General Discussion of Plants:

The AMD treatment system is fully automatic. Ion exchange and filtration are used to produce potable water from acid mine drainage containing excessive amounts of iron and sulfuric acid. The process is essentially one of partial deionization, followed by oxidation, then

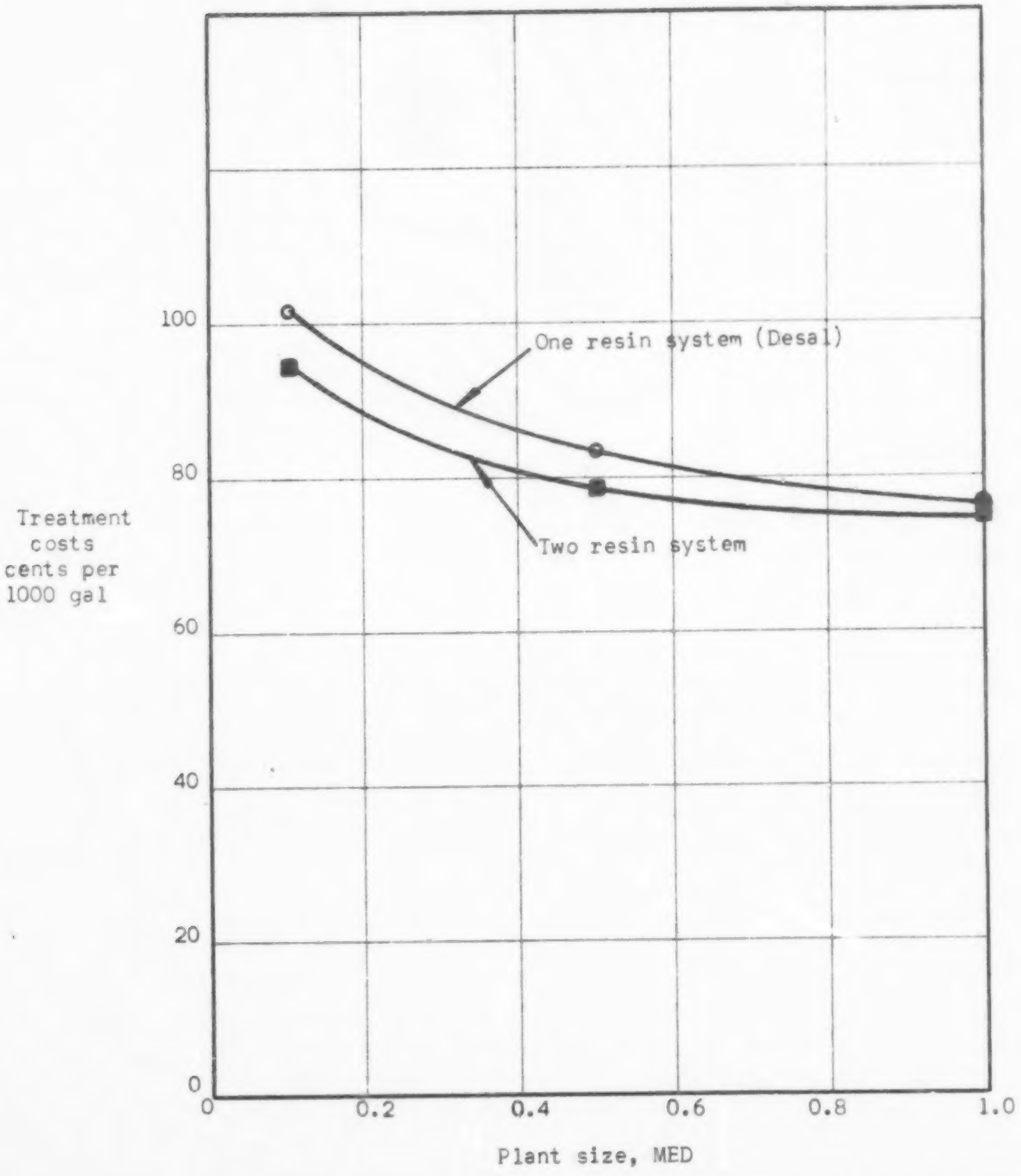

Figure 10 Effect of plant size on treatment costs. Includes equipment and installation labor amortized over ten years, plus chemical regeneration costs. Land, building, labor, utilities, interest costs excluded.

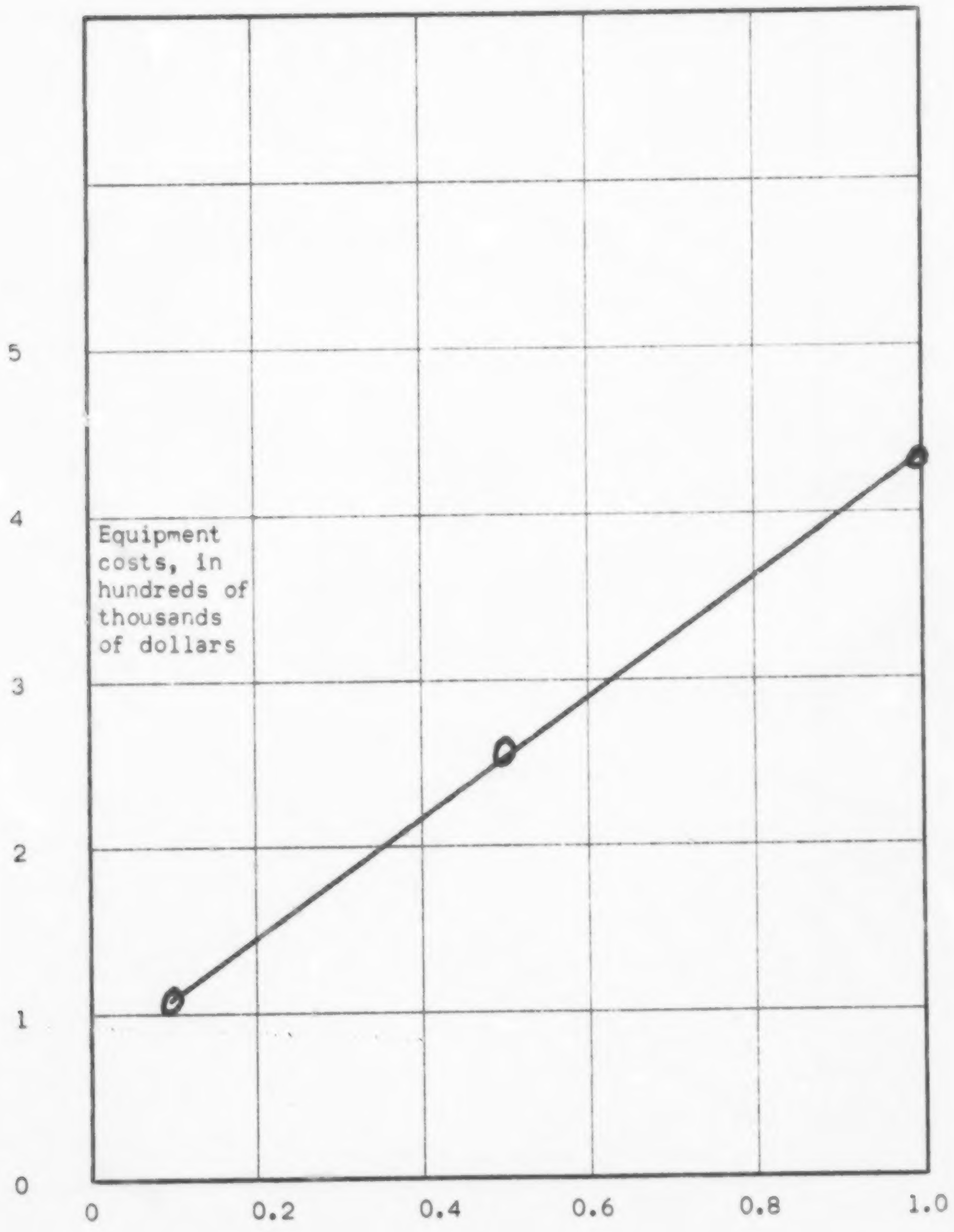

Figure 11. Cost estimates for unassembled, unerected equipment to treat AMD by the Two Resin System.

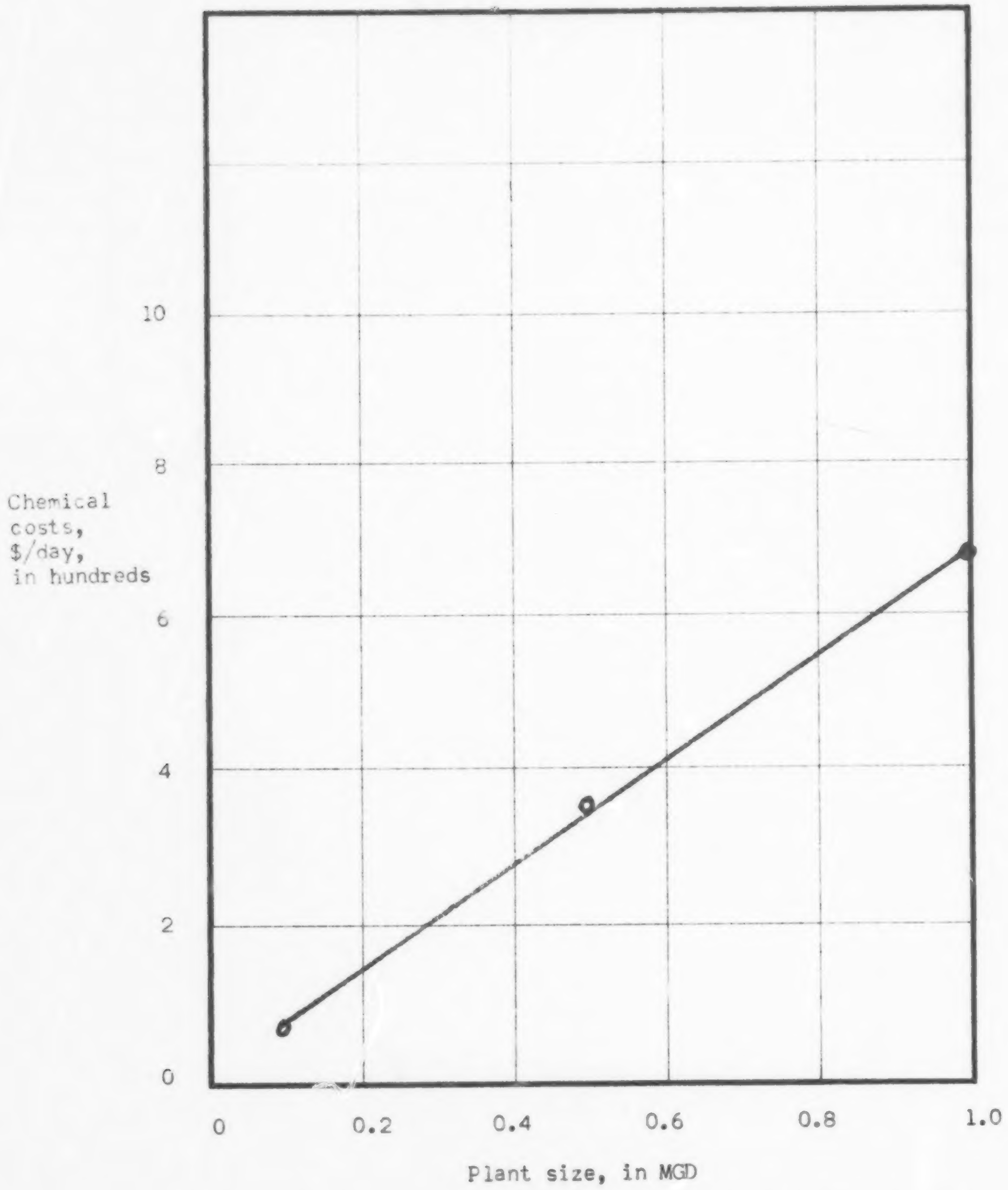

Figure 12 Estimates of daily chemical operating costs to treat AMD by the Two Resin System.

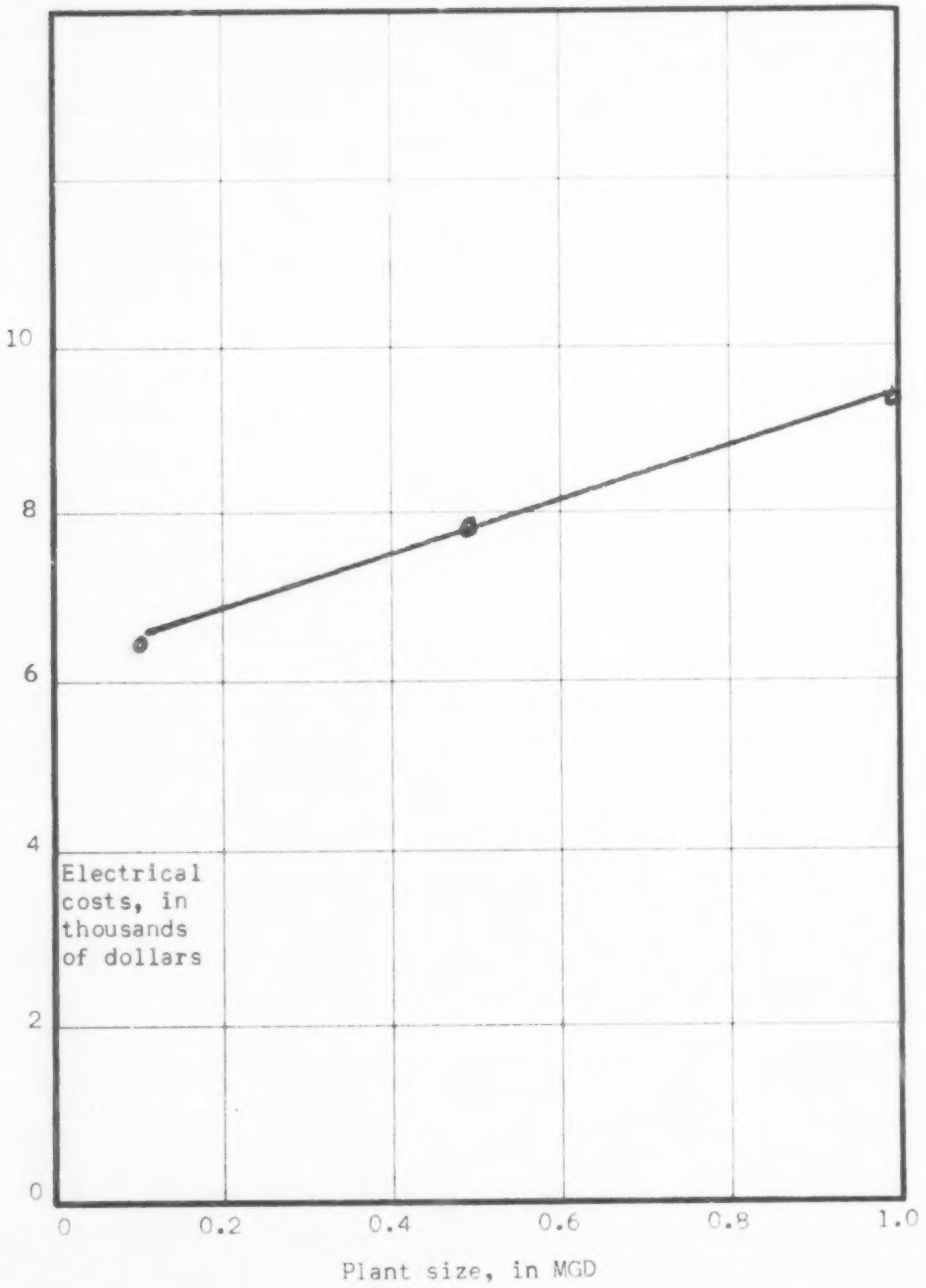

Figure 13 Estimates of electrical labor costs for erection of plants to treat AMD by the Two Resin System.

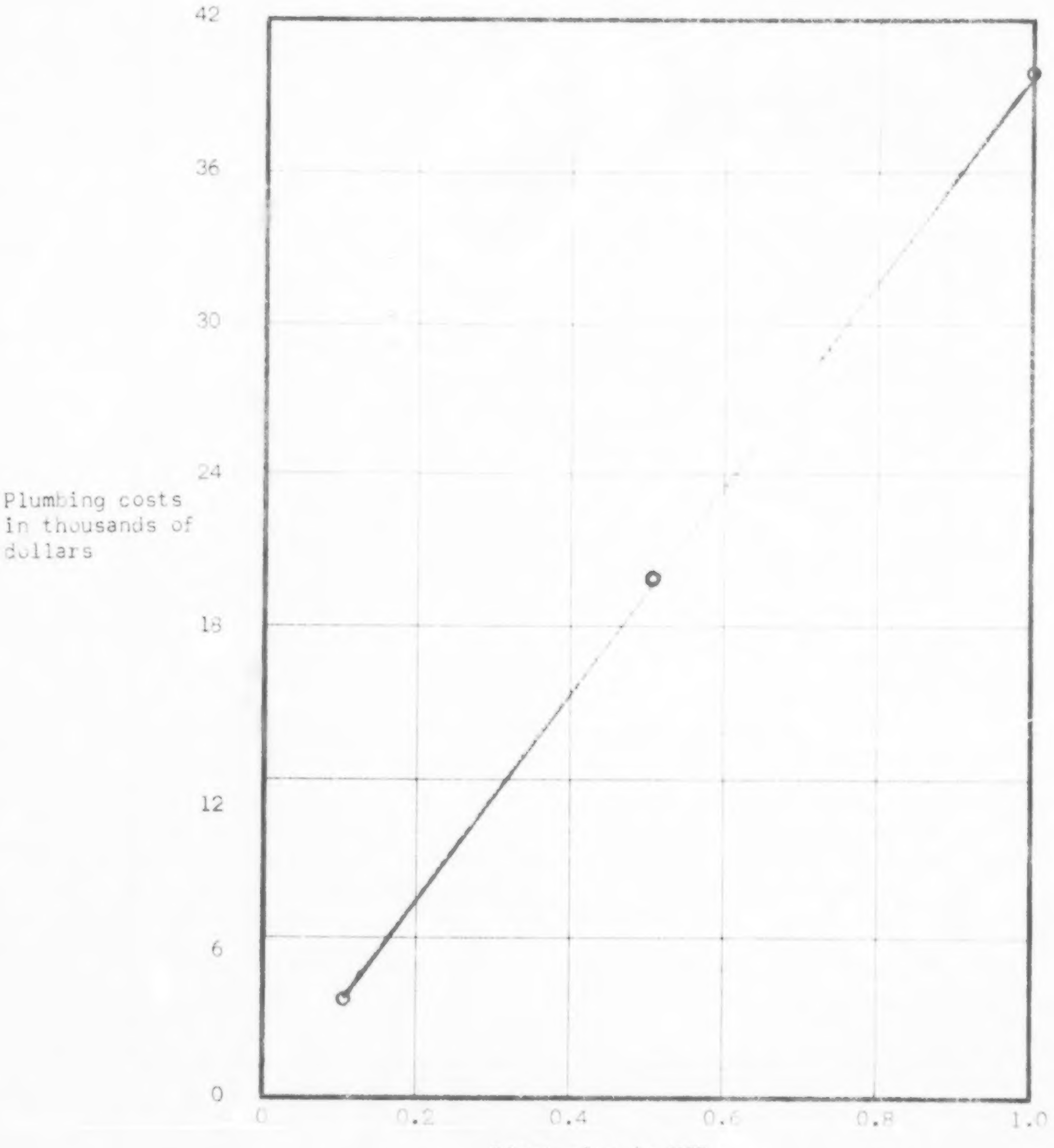

Figure 14 Estimates of plumbing labor costs for assembly and erection of plants to treat AMD by the Two Resin System.

filtration, and finally pH adjustment.

The equipment will consist essentially of two or more pressure vessels (tanks) containing cation exchange resin, two or more tanks containing anion exchange resin, one aerator, and two or more tanks containing granular minerals for final filtration of the product water.

The operation of this treatment system is predicated on the assumption that the AMD will be supplied to the system at a pressure of 75-100 pounds per square inch, will be free of turbidity, and will conform to the raw AMD analysis given elsewhere in this report.

The raw AMD first passes downflow through the cation exchange vessels for removal of most of the polyvalent cations. (Total removal of these cations is not obtained until the process is completed.) Cation leakage occurs because low regeneration levels are used to increase operating efficiency, thereby decreasing the operating costs. Equipment costs are reduced by designing the plant to regenerate each cation (or anion) exchange vessel several times a day.

Each cation and anion exchange vessel will be provided with a conductivity monitor to measure effluent quality. When the conductivity reaches a predetermined value, the monitor will automatically remove that vessel from service and initiate the regeneration process. Several regenerations will occur each day. When an ion exchange vessel is being regenerated, the full service load will be maintained through the units remaining in service. An interlock system will be provided so that not more than one exchange vessel can be regenerated at a time. When regeneration is complete, the vessel will automatically return to service (or standby).

Regeneration of the ion exchange resins will result in a waste which must be treated before disposal. The regenerant effluent strength will be too highly mineralized for reuse. Regenerant rinse time is minimized in view of the high ionic content of the effluent during service. Therefore, no "rinse tailings" are available for possible reuse. While it is not a part of this study or plant design, all of the backwash wastes, regenerant wastes, and rinse water should be collected in a common waste lagoon.

During regeneration, the ion exchange resins must first be backwashed. This will remove insolubles which have been physically removed by filtration during service. The cation exchange resin will be backwashed with acid mine drainage.

The cation exchange resins will be regenerated downflow with 2% sulfuric acid. An acid-resistant pump will transfer 66° Be sulfuric acid from the bulk storage facilities. Dilution water will be AMD. A conductivity meter will continuously indicate the percent strength of dilute regenerant acid influent to the vessel during regeneration. Regenerant rinse water will be AMD. The regeneration steps will all be time controlled. At the completion of rinse, the regenerated cation exchange vessel will return to service automatically.

In the second step of the AMD treatment process, the effluent from the cation exchanger passes downflow through the multiple anion exchange vessels where the mineral acids are removed. In addition to treating water for product, these anion exchangers will supply rinse water for the final filters.

The anion exchange resin will be regenerated downflow with 4% caustic soda (sodium hydroxide). A caustic resistant pump will transfer 50% caustic solution from the bulk storage facilities. The caustic will be diluted to the desired 4% solution with the treated effluent from the cation exchange vessels. A conductivity indicator will indicate the percent concentration of the regenerant as it is introduced to the anion exchanger. The regeneration will be time controlled. At the completion of the rinse, the regenerated anion exchanger will return to service or standby automatically.

The product from the anion exchangers still contains objectionable amounts of manganese. This is precipitated by raising the pH to about 10 and aerating. The pH adjustment uses a chemical feed pump to add sodium hydroxide solution. The feed rate is pH controlled. Aeration is in a standard tray-type forced draft aerator, with air blown upward to intimately contact the downward flow of water. This counterflow design assures oxidation of iron and manganese, which precipitate at the elevated pH. This aerator, rather than surface aeration, is selected for its greater efficiency.

Water dropping through the aerator is collected in a reservoir, sized to retain about five minutes flow. The reservoir float control will regulate the inlet valve to the aerator, thereby maintaining the water level in the reservoir. This level will be subject to significant changes only when the filters are backwashed once a day.

A pump transfers the aerated water from the reservoir to the multiple high velocity filters for final filtration. The filtered water will require pH reduction to meet the requirements of potable water. To accomplish this, a small amount of treated water from the cation exchangers will be blended with filtered effluent. A pH recorder/controller will control the cation exchanger effluent to maintain the proper final pH level.

The filters will be backwashed with treated water from the cation exchangers. The conditioning rinse, prior to returning to service is from the aerator clear well (as in service). The backwash and rinse effluents will be sent to the waste lagoon. The backwash and rinse operations will be time controlled, automatically initiated once a day for each filter. Filters are so sized that removal of one filter from service will not overload the system, nor interrupt the product flow of water.

Materials of construction were chosen in accordance with accepted standards for the ion exchange industry. PVC pipe has not been used because of inherent physical weakness in the pipe sizes used. Stainless steel pipe has not been used because of its excessive cost, compared to saran lined steel. The latter has been proven to be thoroughly acceptable in such applications. Concentrated sulfuric acid is shipped and stored in unlined steel tanks: moisture must be avoided during storage; therefore, air vents must include a desiccant. Caustic soda solutions are stored in unlined steel containers - corrosion is negligible.

Individual Plant Specifications:

Plant to produce 0.1 MGD:

The equipment will consist essentially of two cation exchange vessels, two anion exchange vessels, one aerator, and two final filters.

The plant is designed to produce a nominal 100,000 gallons per day of water with mineral content not exceeding that specified for potable water. It is anticipated that there will be 20,900 gpd waste solutions. This volume will contain approximately 620 pounds of 100% sulfuric acid. The treatment and disposal of this material is not included in this study, but consideration of this problem should be in the overall concept of the project, and may be a major cost item.

The detailed specifications on the equipment for this 0.1 MGD plant are:

Table 30	Hydrogen Cation Exchangers (Regenerations total 6/day)
Table 31	Anion Exchangers (Regenerations total 6/day)
Table 32	Forced Draft Degasifier or Aerator
Table 33	Pressure Filters
Table 34	Miscellaneous Items and Exclusions
Figure 15	Flow Diagram
Figure 16	Plant Plan

Chemical Operation costs are estimated as follows:

Sulfuric acid, 66° Be, 1,728 lb/day, $0.016/lb = $27.65
Caustic soda, 987 lb/day, $0.37/lb = 36.52
Total, $/day = $64.17

Cost estimates for the equipment as specified in Tables 30 through 34 for this 0.1 MGD size plant, excluding freight, building and land, and assembly and erection, total $106,000.

Cost estimates for erection have been made as follows:

Electrical	$6,200
Plumbing	2,000

TABLE 30
HYDROGEN CATION EXCHANGERS
DETAILED SPECIFICATIONS, 0.1 MGD

PERFORMANCE:

Total System

Total influent cations, gpg as $CaCO_3$	63.7
Design flow rate, gpm	70
Operating water pressure, psig	75
Number of units	Two

Per Unit

Design flow rate, gpm	70
Peak flow rate, gpm	120
Backwash rate, gpm	78
Cation exchange material, type	Strongly Acidic
quantity, cu. ft.	96
capacity, Kgr per cu. ft.	12.5
capacity, Kgr per unit	1,200
Gallons treated per regeneration (includes anion regeneration water)	18,800
Gallons treated to service (net)	16,785
Regenerant, type	H_2SO_4 -66° Be
quantity per regeneration	288 lbs

SPECIFICATIONS:

Model number	None

Tanks

Tank diameter	60"
Straight side of tank	108"
Design working pressure of tank	100 psi Non-Code
External surface	Prime Painted
Tank lining, material and thickness	3/32" Plasticol
Tank supports	Adjustable Jacks
Access opening(s)	12" x 16" Manhole

Internals

Inlet distributor, design and materials	PVC Header Lateral
Regenerant distributor, design and materials	PVC Header Lateral
Underdrain system, design and materials	PVC Header Lateral
Supporting bed	Silica Gravel

Piping

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Main piping size	3"
Main piping material	Saran Lined Steel
Main valving arrangement	Nest of Diaphragm Type
Main valving material	Saran Lined Cast Iron

Table continued on next page

TABLE 30A
HYDROGEN CATION EXCHANGERS continued
DETAILED SPECIFICATIONS, 0.1 MGD

Control System

Control	Automatic
Initiation of regeneration	Conductance Ratio Meter
Backwash control	Limit Stop On Valve

Auxiliaries

Meter, size and type	None
Meter register	-
Interconnecting piping between multiple units, inlet & outlet	3" Saran Lined
Pressure gauges	Chemical Seal Type
Sample cocks	1 Pair Per Unit

Regeneration Equipment

Type of regenerant introduction	Pump
Regenerant introduction strength	2%
Regenerant tank size bulk storage	5 Ft Dia x 8 Ft
Material of construction	Unlined Steel

Electrical Requirements

Volts, Hertz, Phase	115/60/1

ADDITIONAL SPECIFICATIONS:

Conductance ratio bridges to determine end of cycle	2 - Model RE-18G
Special sampling manifolds for ratio bridges	Screened Header-Lateral-PVC
Solu-Bridge to monitor acid regenerant strength with selector switch and two cells	Model RD-226C
Bypass type rate of flow meter on each unit	Included
Horizontal acid storage tank for concrete saddles	Included
Regenerant acid piping system	1/2" Carpenter 20 SS
Waste discharge inter-connecting heads	3" Saran Lined
Special Milton Roy Acid Pump, 1/2 HP - T. E. Motor	34 gph

TABLE 31

ANION EXCHANGERS

DETAILED SPECIFICATIONS, 0.1 MGD

PERFORMANCE:

Total System

Total influent exchangeable anions, gpg as $CaCO_3$	84.0
Design flow rate, gpm	70
Operating water pressure, psig	50
Number of units	Two

Per Unit

Design flow rate, gpm	70
Peak flow rate, gpm	70
Backwash rate, gpm	50
Anion exchange material, type	Weakly Basic
quantity, cu. ft.	47
capacity, Kgr per cu. ft.	30
capacity, Kgr per unit	1,410
Gallons treated per regeneration	16,785
Regenerant, type	NaOH-50% Liquid
quantity per regeneration	25.8 gal (165 lb, 100%)

SPECIFICATIONS:

Model Number	None

Tanks

Tank diameter	48"
Straight side of tank	96"
Design working pressure of tank	100 psi Non-Code
External surface	Prime Painted
Tank lining, material and thickness	3/32" Plastisol
Tank supports	Adjustable Jacks
Access opening(s)	12" x 16" Manhole

Internals

Inlet distributor, design and materials	PVC Header Lateral
Regenerant distributor, design and materials	PVC Header Lateral
Underdrain system, design and materials	PVC Header Lateral
Supporting bed	Silica Gravel

Piping

Main piping size	3"
Main piping material	Saran Lined Steel
Main valving arrangement	Nest of Diaphragm Type
Main valving material	Saran Lined Cast Iron

Table continued on next page

TABLE 31A

ANION EXCHANGERS continued
DETAILED SPECIFICATIONS, 0.1 MGD

Control System

Control	Automatic
Initiation of regeneration	Conductance Ratio Meter
Backwash control	Limit Stop on Valve

Auxiliaries

Meter, size and type	None
Meter register	-
Interconnecting piping between multiple units	Included
Pressure gauges (inlet & outlet)	Chemical Seal Type
Sample cocks	1 Pair Per Unit
Conductivity instrument, type	Conductance Ratio
manufacturer	Beckman
model number	RE-18G

Regeneration Equipment

Type of regenerant introduction	Pump
Regenerant introduction strength	4%
Regenerant tank size, inches, bulk storage	5 ft dia x 11 ft
Material of construction	Unlined Steel

Electrical Requirements

Volts, Hertz, Phase	115/60/1

ADDITIONAL SPECIFICATIONS:

Special sampling manifolds for ratio bridges	Screened. PVC Header lateral
Solu-Bridge to monitor caustic regenerant strength with 2 cells and selector switch	Model RD-227C
Bypass rate of flow meter on each unit	Included
Caustic storage tank for concrete saddles	Included
Regenerant caustic piping system	3/4" Wrought Steel
Special Milton Roy Caustic Pump, 1/2 HP T. E. Motor	78 gph
Inlet headers for cation effluent	3" Saran Lined
Outlet header for anion effluent	3" Wrought Steel
Waste discharge inter-connecting headers	3" Saran Lined

TABLE 32

FORCED DRAFT DEGASIFIERS OR AERATOR

DETAILED SPECIFICATIONS, 0.1 MGD

PERFORMANCE:

Total System

Function	Aeration
Influent iron & manganese content, ppm	2.4
Influent temperature, °F	Ambient
Design flow rate, gpm	70
Water pressure at inlet	15 psi
Number of units	One

Per Unit

Design flow rate, gpm	70
Peak flow rate, gpm	70

SPECIFICATIONS:

Tower

Size of tower	36"
Height of tower	144"
Materials of construction	Fir Staves

Internals

Material of packing	Redwood Trays
Depth of packing	9 ft
Inlet distributor, design and materials	PVC Header Lateral
Support, design and materials	Wood

Auxiliaries

Blower, type	Centrifugal
capacity, cfm	840
static head, inches H_2O	2"
Motor, type	1 HP - ODP
voltage, current, phases	230-460/60/3
Level control, type - for reservoir	Modulating "Leveltrol"
Inlet valve, type	Modulated
material of construction	Cast Iron, SS Trim
size	1-1/2"

Storage

Concrete reservoir below floor level	Not Included
Low water float switch for pump safety cutoff	Included
Forwarding pump, 70 gpm @ 100-120 TDH, 7.5 HP-ODP Motor, all stainless construction	Included

TABLE 33

PRESSURE FILTERS
DETAILED SPECIFICATIONS, 0.1 MGD

PERFORMANCE:	
Total System	
Design flow rate, gpm	70
Operating water pressure, psig	30 minimum
Number of units	Two
Type of units	Hi-Velocity
Per Unit	
Design flow rate, gpm	35
Peak flow rate, gpm	50
Design rate, gpm/ft^2 of filter area	7.1
Backwash rate, gpm	73
Filter media, type	Hi-Velocity
quantity, cu. ft.	13.5
depth of bed, inches	33
SPECIFICATIONS:	
Model Number	None
Tanks	
Tank diameter	30"
Straight side of tank	60"
Design working pressure of tank	100 psi Non-Code
External surface	Prime Painted
Internal surface	3/32" Plastisol
Tank supports	Adjustable Jacks
Access opening(s)	12" x 16" Manhole
Internals	
Inlet distributor, design and materials	PVC Header Lateral
Underdrain system, design and materials	PVC Header Lateral
Supporting bed	Gravel
Piping	
Main piping size	2"
Main piping material	Saran Lined
Main valving arrangement	Nest of Diaphragm Valves
Control System	
Control	Automatic
Initiation of regeneration	Timeclock
Backwash control	Limit Stops on Valves

Table continued on next page

TABLE 33A

PRESSURE FILTERS (Cont'd)

Auxiliaries	
Interconnecting piping between multiple units	2" Wrought Steel
Pressure gauges	2 Pair Included
Sample cocks	2 Pair Included
Electrical Requirements	
Voltz	115
Hertz	60
Phase	1

ADDITIONAL SPECIFICATIONS:

Bypass type rate of flow meter, two on each unit	Included
Waste Discharge inter-connecting headers	3" PVC

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TABLE 34
MISCELLANEOUS ITEMS INCLUDED:

Detailed Specifications, 0.1 MGD

Air compressor for all plant control needs
Modulating pH meter, alarm, pH flow cells and electrodes with milliamp output to control pH adjustment pump
Modulating pH meter as above but with transducer added for 3-15 psi air signal to control final acid blending - neutralizing valve
Two-pen strip chart records for pH instruments
Alloy 20 blending valve for cation effluent, which controls to modulate feed from 3-15 psi air signal
Caustic feed pump to adjust pH prior to aeration, equipped with controller station to modulate feed from milliamp input
All inter-plant piping needed to inter-connect ion exchange units, aerator and filters with saran lining steel, and black wrought steel, and PVC, of various sizes

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The following items are specifically excluded:

1. The building, its foundations, concrete reservoir below aerator
2. Auxiliary plumbing and plumbing fixtures for the building
3. Electrical wiring of building and connections between electrical controls: **costs separately estimated**
4. All pump starters: **costs separately estimated**

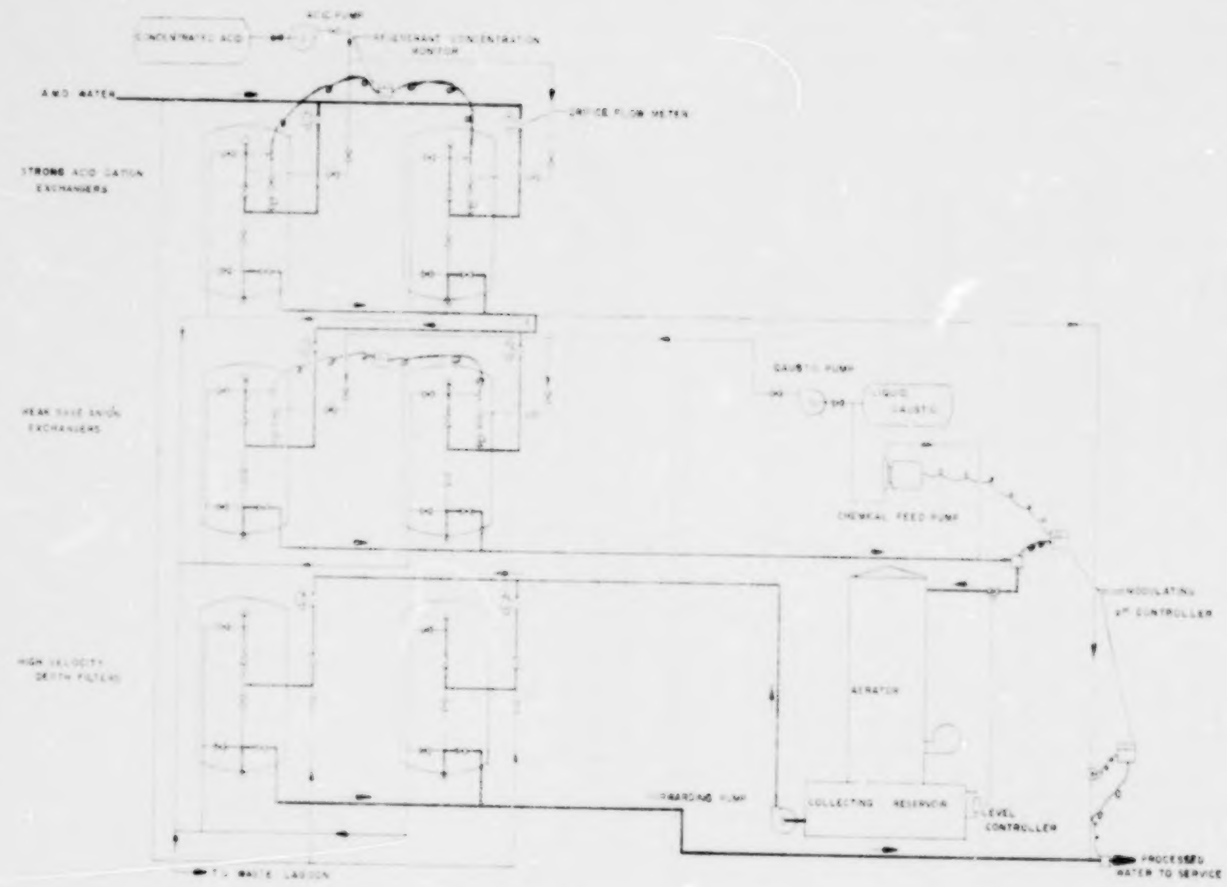
CONCENTRATED ACID
ACID PUMP
REGENERANT CONCENTRATION MONITOR
AMD WATER
ORIFICE FLOW METER
STRONG ACID CATION EXCHANGERS
WEAK BASE ANION EXCHANGERS
HIGH VELOCITY DEPTH FILTERS
CAUSTIC PUMP
LIQUID CAUSTIC
CHEMICAL FEED PUMP
MODULATING pH CONTROLLER
AERATOR
FORWARDING PUMP
COLLECTING RESERVOIR
LEVEL CONTROLLER
PROCESSED WATER TO SERVICE
TO WASTE LAGOON

FIGURE 15. AMD Treatment Plant Flow Diagram, 2-Resin System 0.1 MGD

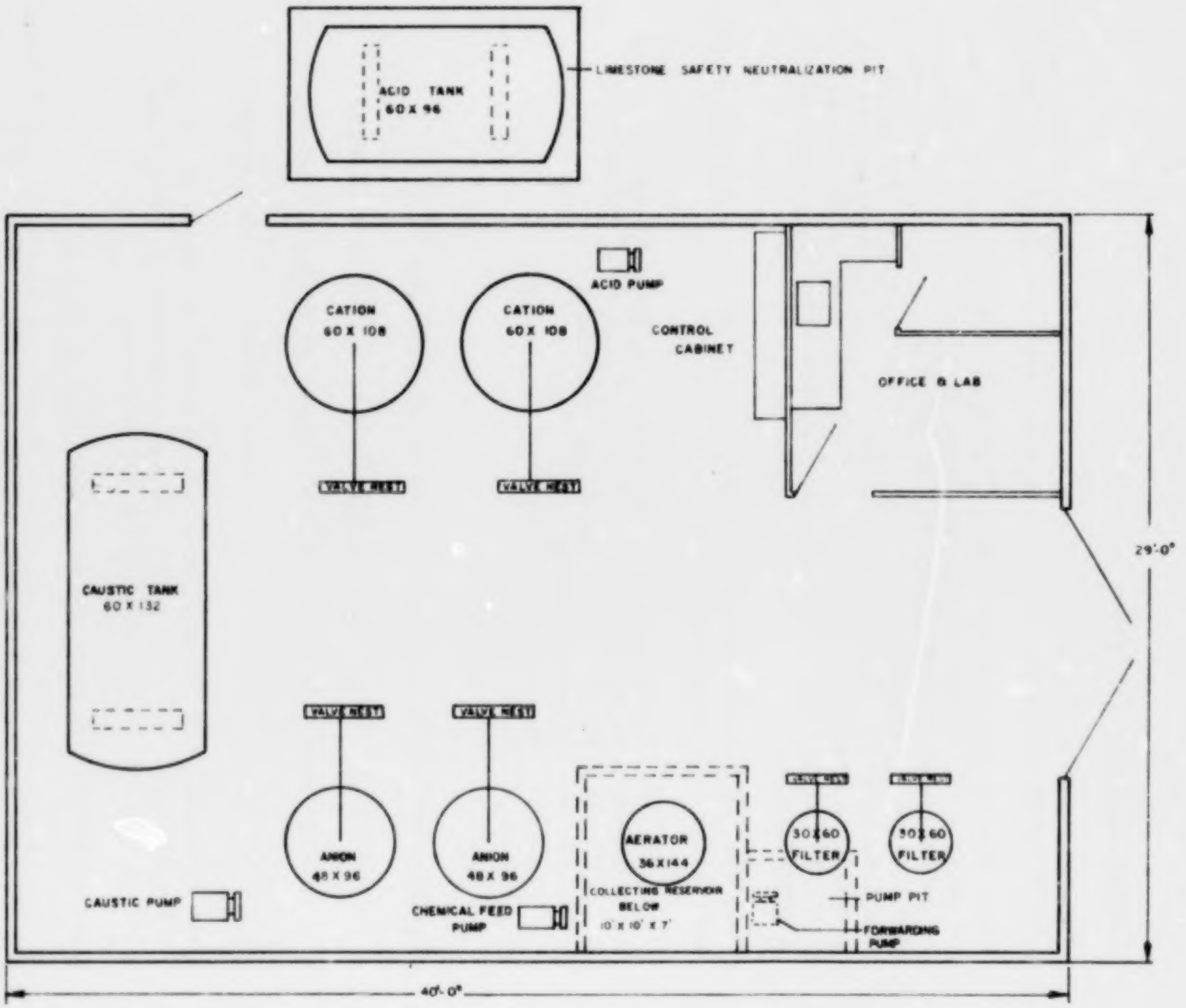
ACID TANK
60 X 96
LIMESTONE SAFETY NEUTRALIZATION PIT
ACID PUMP
CATION
60 X 108
CATION
60 X 108
CONTROL CABINET
OFFICE & LAB
VALVE NEST
VALVE NEST
CAUSTIC TANK
60 X 132
29'-0"
VALVE NEST
VALVE NEST
ANION
48 X 96
ANION
48 X 96
AERATOR
36 X 144
COLLECTING RESERVOIR BELOW
10' X 10' X 7'
30 X 60
FILTER
30 X 60
FILTER
PUMP PIT
FORWARDING PUMP
CAUSTIC PUMP
CHEMICAL FEED PUMP
40'-0"

FIGURE 16. AMD Treatment Plant Plan, 2-Resin system, 0.1 MGD

Plant to produce 0.5 MGD:

The equipment will consist essentially of two cation exchange vessels, two anion exchange vessels, one aerator, and three final filters.

The plant is designed to produce a nominal 500,000 gallons per day of water with mineral content not exceeding that for potable water. It is anticipated that there will be 106,000 gallons of waste per day containing approximately 3,000 pounds of 100% sulfuric acid. The treatment and disposal of this material is not included in this study, but consideration of the problem should be in the overall concept of the project. This problem investigation is included in the recommendations.

The detailed specifications of the equipment for this 0.5 MGD plant are on the following pages:

Table 35	Hydrogen Cation Exchangers (Regenerations total 6/day)
Table 36	Anion Exchangers (Regenerations total 6/day)
Table 37	Forced Draft Degasifier or Aerator
Table 38	Pressure Filters
Table 39	Miscellaneous Items and Exclusions
Figure 17	Flow Diagram
Figure 18	Plant Plan

Chemical operating costs are estimated as follows:

Sulfuric acid, 66° Be,	8,532 lb/day, $0.016/lb	= $136.51
Caustic soda, 50%,	4,914 lb/day, $0.037/lb	= 181.82
	Total $/day	= $318.33

Cost estimates for the equipment as specified in Tables 35 through 39 for this 0.5 MGD size plant, excluding freight, building and land, and assembly and erection total $256,000.

Cost estimates for erection have been made as follows:

Electrical	$ 7,800
Plumbing	20,000

TABLE 35

HYDROGEN CATION EXCHANGERS

DETAILED SPECIFICATIONS, 0.5 MGD

PERFORMANCE:

Total System

Total influent cations, gpg as $CaCO_3$	63.7
Design flow rate, gpm	351
Operating water pressure, psig	75
Number of units	Two

Per Unit

Design flow rate, gpm	351
Peak flow rate, gpm	605
Backwash rate, gpm	313
Cation exchange material, type	Strongly Acidic
quantity, cu. ft.	474
capacity, Kgr per cu. ft.	12.5
capacity, Kgr per unit	5,925
Gallons treated per regeneration (includes anion regeneration water)	92,900
Gallons treated to service (net)	83,500
Regenerant, type	H_2SO_4 -66° Be
quantity per regeneration	1,422 lbs

SPECIFICATIONS:

Model number	None

Tanks

Tank diameter	120"
Straight side of tank	120"
Design working pressure of tank	100 psi Non-Code
External surface	Prime Painted
Tank lining, material and thickness	3/32" Plasticol
Tank supports	Adjustable jacks
Access opening(s)	12" X 16" Manhole

Internals

Inlet distributor, design and materials	PVC header lateral
Regenerant distributor, design and materials	PVC " "
Underdrain system, design and materials	" " "
Supporting bed	Silica Gravel

Piping

Main piping size	6"
Main piping material	Saran Lined Steel
Main valving arrangement	Nest Of Diaphragm Type
Main valving material	Saran Lined Cast Iron

Table continued on next page

TABLE 35A

HYDROGEN CATION EXCHANGERS continued

Control System Detailed Specifications, 0.5 MGD

Control	Automatic
Initiation of regeneration	Conductance Ratio Meter
Backwash control	Limit Stop On Valve
Auxiliaries	
Meter, size and type	None
Meter register	-
Interconnecting piping between multiple units, inlet & outlet	6" Saran Lined
Pressure gauges	Chemical Seal Type
Sample cocks	1 pair per unit
Regeneration Equipment	
Type of regenerant introduction	Pump
Regenerant introduction strength	2%
Regenerant tank size, bulk storage	8 ft diam X 15 ft
Material of construction	Unlined Steel
Electrical Requirements	
Volts, Hertz, Phase	115/60/1

ADDITIONAL SPECIFICATIONS:

Conductance ratio bridges to determine end of cycle	2 -Model RE-18G
Special sampling manifolds for ratio bridges	Screened header-lateral-PVC
Solu-Bridge to monitor acid regenerant strength with selector switch and 2 cells	Model RD-226C
Bypass type rate of flow meter on each unit	Included
Horizontal acid storage tank for concrete saddles	Included
Regenerant acid piping system	1/2" Carpenter 20 SS
Waste discharge inter-connecting heads	6" Saran Lined
Special Milton Roy Acid Pump, 3/4 HP - T. E. Motor	168 gph

TABLE 36

ANION EXCHANGERS

Detailed Specifications, 0.5 MGD

PERFORMANCE:

Total System

Total influent exchangeable anions, gpg as $CaCO_3$	84.0
Design flow rate, gpm	351
Operating water pressure, psig	50
Number of units	Two

Per Unit

Design flow rate, gpm	351
Peak flow rate, gpm	351
Backwash rate, gpm	254
Anion exchange material, type	Weakly Basic
quantity, cu. ft.	234
capacity, Kgr per cu. ft.	30
capacity, Kgr per unit	7,014
Gallons treated per regeneration	83,500
Regenerant, type	NaOH-50% liquid
quantity per regeneration	128.4 gal (819 lb, 100%)

SPECIFICATIONS:

Model Number	None

Tanks

Tank diameter	108"
Straight side of tank	96"
Design working pressure of tank	100 psi non-Code
External surface	Prime painted
Tank lining, material and thickness	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" X 16" Manhole

Internals

Inlet distributor, design and materials	PVC header lateral
Regenerant distributor, design and materials	" " "
Underdrain system, design and materials	" " "
Supporting bed	Silica Gravel

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Piping

Main piping size	6"
Main piping material	Saran lined Steel
Main valving arrangement	Nest of Diaphragm type
Main valving material	Saran lined cast iron

Table continued on next page

TABLE 36A

ANION EXCHANGERS continued
Detailed Specifications, 0.5 MGD

Control System

Control	Automatic
Initiation of regeneration	Conductance Ratio Meter
Backwash control	Limit stop on valve

Auxiliaries

Meter, size and type	None
Meter register	-
Interconnecting piping between multiple units (inlet & outlet)	6" included
Pressure gauges	Chemical seal type
Sample cocks	1 pair per unit
Conductivity instrument, type	Conductance Ratio
manufacturer	Beckman
model number	RE-18G

Regeneration Equipment

Type of regenerant introduction	Pump
Regenerant introduction strength	4%
Regenerant tank size, bulk storage	8 ft diam X 20 ft
Material of construction	Unlined steel

Electrical Requirements

Volts, Hertz, Phase	115/60/1 & 230-460/60/3

ADDITIONAL SPECIFICATIONS:

Special sampling manifolds for ratio bridges	Screened, PVC header lateral
Solu-Bridge to monitor caustic regenerant strength with 2 cells and selector switch	Model RD-227C
Bypass rate of flow meter on each unit	Included
Caustic storage tank for concrete saddles	Included
Regenerant caustic piping system	3/4" wrought steel
Special Milton Roy caustic pump, 2 HP T.E. Motor	384 gph
Inlet headers for cation effluent	6" Saran lined
Outlet header for Anion effluent	6" wrought steel
Waste discharge inter-connecting headers	6" Saran lined

TABLE 37

FORCED DRAFT DEGASIFIERS OR AERATOR

Detailed Specifications, 0.5 MGD

PERFORMANCE:

Total System

Function	Aeration
Influent iron & manganese content, ppm	2.4
Influent temperature, °F.	Ambient
Design flow rate, gpm	351
Water pressure at inlet	15 psi
Number of units	One

Per Unit

Design flow rate, gpm	351
Peak flow rate, gpm	351

SPECIFICATIONS:

Tower

Size of tower	72"
Height of tower	144"
Materials of construction	Fir Staves

Internals

Material of packing	Redwood trays
Depth of packing	9 ft
Inlet distributor, design and materials	PVC header lateral
Support, design and materials	Wood

Auxiliaries

Blower, type, Two Units Needed	Centrifugal
capacity, cfm	2100 each
static head, inches H_2O	2"
Motor, type	1.5 HP - ODP
voltage, current, phases	230-460/60/3
Level control, type - for reservoir	Modulating "Leveltrol"
Inlet valve, type	Modulated
material of construction	Cast iron, SS trim
size	3"

Storage

Concrete reservoir below floor level	Not included
Low water float switch for pump safety cutoff	Included
Forwarding pump, 351 gpm @ 100-120 TDH, 20 HP -ODP Motor, all stainless construction	Included

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TABLE 38

PRESSURE FILTERS
Detailed Specifications, 0.5 MGD

PERFORMANCE:	
Total System	
Design flow rate, gpm	351
Operating water pressure, psig	30 minimum
Number of units	Three
Type of units	Hi-Velocity
Per Unit	
Design flow rate, gpm	117
Peak flow rate, gpm	117
Design rate, gpm/ft^2 of filter area	9.4
Backwash rate, gpm	190
Filter media, type	Hi-Velocity
quantity, cu. ft.	35
depth of bed, inches	33
SPECIFICATIONS:	
Model Number	None
Tanks	
Tank diameter	48"
Straight side of tank	72"
Design working pressure of tank	100 psi non-code
External surface	Prime painted
Internal surface	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" X 16" Manhole
Internals	
Inlet distributor, design and materials	PVC Header lateral
Underdrain system, design and materials	" " "
Supporting bed	Gravel
Piping	
Main piping size	3" X 4"
Main piping material	Saran Lined
Main valving arrangement	Nest of diaphragm valves
Control System	
Control	Automatic
Initiation of regeneration	Timeclock
Backwash control	Limit stops on valves

Table continued on next page

TABLE 38A
PRESSURE FILTERS (Cont'd)

Auxiliaries	
Interconnecting piping between multiple units, inlet & outlet	6" wrought steel
Pressure gauges	3 pair included
Sample cocks	3 " "
Electrical Requirements	
Voltz	115
Hertz	60
Phase	1

ADDITIONAL SPECIFICATIONS:

Bypass type rate of flow meter	2 per unit
Waste discharge inter-connecting headers	6" Saran lined

TABLE 39
MISCELLANEOUS ITEMS INCLUDED:

Detailed Specifications, 0.5 MGD

Air compressor for all plant control needs
Modulating pH meter, alarm, pH flow cells and electrodes with milliamp output to control pH adjustment pump
Modulating pH meter as above but with transducer added for 3-15 psi air signal to control final acid blending - neutralizing valve
Two-pen strip chart records for pH instruments
Alloy 20 blending valve for cation effluent, which controls to modulate feed from 3-15 psi air signal
Caustic feed pump to adjust pH prior to airation, equipped with controller station to modulate feed from milliamp input
All inter-plant piping needed to inter-connect ion exchange units, aerator and filters with Saran lining steel, and black wrought steel, and PVC, of various sizes

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The following items are specifically excluded:

1. The building, its foundations, concrete reservoir below aerator
2. Auxilliary plumbing and plumbing fixtures for the building
3. Electrical wiring of building and connections between electrical controls: Costs separately estimated
4. All pump starters: Costs separately estimated

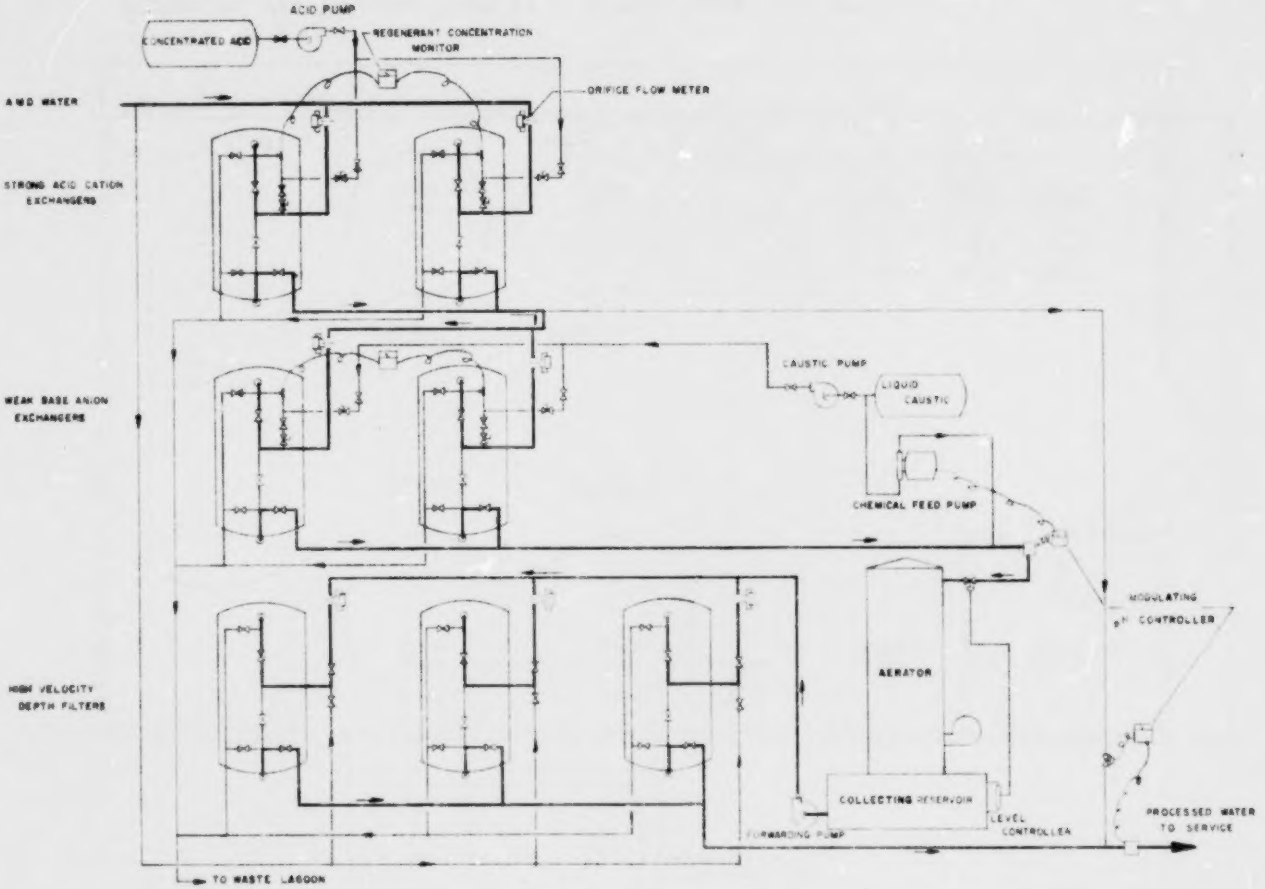
CONCENTRATED ACID
ACID PUMP
REGENERANT CONCENTRATION MONITOR
ORIFICE FLOW METER
AMD WATER
STRONG ACID CATION EXCHANGERS
WEAK BASE ANION EXCHANGERS
CAUSTIC PUMP
LIQUID CAUSTIC
CHEMICAL FEED PUMP
MODULATING pH CONTROLLER
HIGH VELOCITY DEPTH FILTERS
AERATOR
COLLECTING RESERVOIR
FORWARDING PUMP
LEVEL CONTROLLER
PROCESSED WATER TO SERVICE
TO WASTE LAGOON

FIGURE 17. AMD Treatment Plant Flow Diagram, 2-Resin System, 0.5 MGD

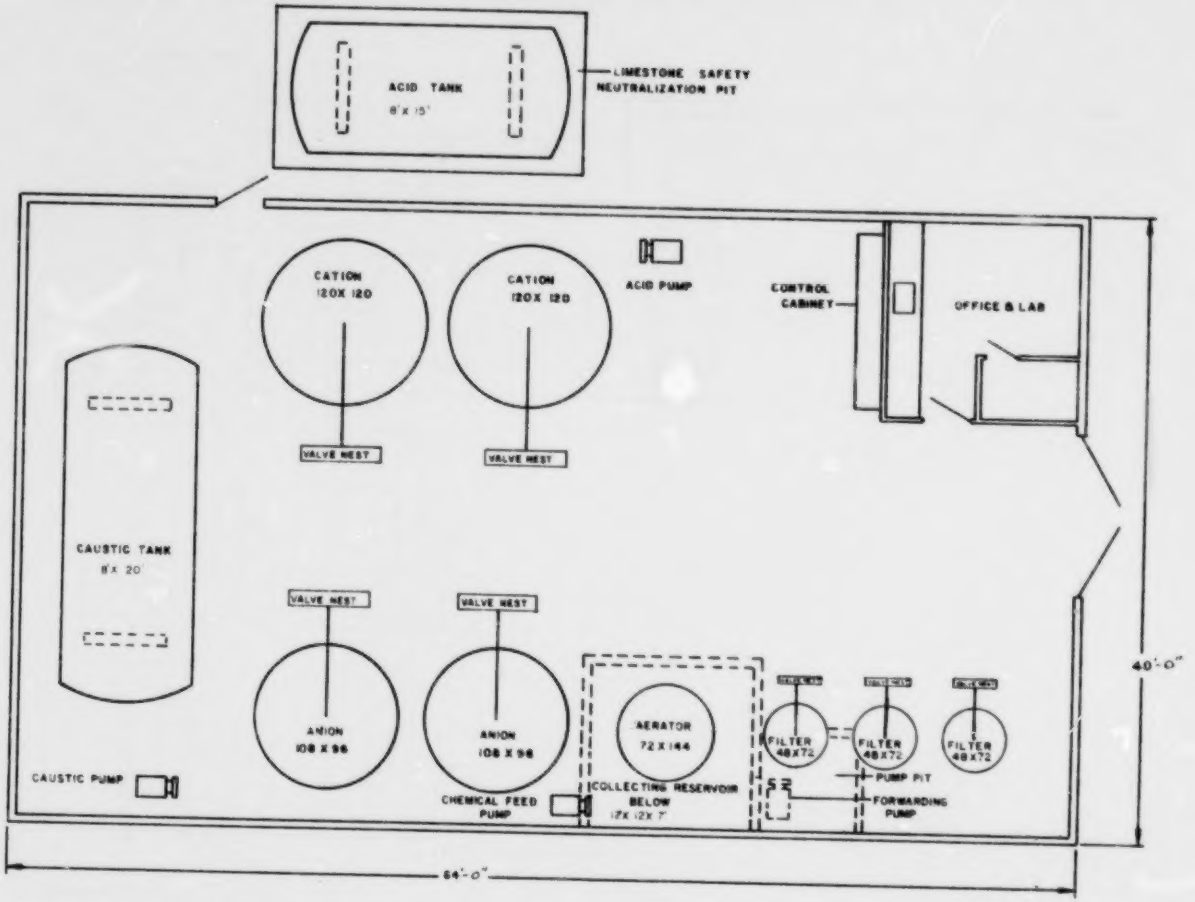
ACID TANK
8'x 15'
LIMESTONE SAFETY NEUTRALIZATION PIT
CATION
120X 120
CATION
120X 120
ACID PUMP
CONTROL CABINET
OFFICE & LAB
VALVE NEST
VALVE NEST
CAUSTIC TANK
8'x 20
VALVE NEST
VALVE NEST
ANION
108 X 96
ANION
108 X 96
AERATOR
72 X 144
COLLECTING RESERVOIR BELOW
12X 12X 7'
FILTER
48X72
FILTER
48X72
FILTER
48X72
PUMP PIT
FORWARDING PUMP
CAUSTIC PUMP
CHEMICAL FEED PUMP
40'-0"
64'-0"

FIGURE 18. AMD Treatment Plant Plan, 2 Resin System, 0.5 MGD

Plant to produce 1.0 MGD:

The equipment will consist essentially of four cation exchange vessels. three anion exchange vessels, one aerator and three final filters.

The plant is designed to produce a nominal 1,000,000 gallons per day of water with mineral content not exceeding that specified for potable water. It is anticipated that there will be 200,000 gallons of waste per day containing approximately 6,300 pounds of 100% sulfuric acid. The treatment and disposal of this material is not included in this study, but consideration of the problem should be in the overall concept of the project.

The detailed specifications of the equipment for this plant are on the following pages:

Table 40	Hydrogen Cation Exchangers (Regenerations total 12/day)
Table 41	Anion Exchangers (Regenerations total 9/day)
Table 42	Forced Draft Degasifier or Aerator
Table 43	Pressure Filters
Table 44	Miscellaneous Items and Exclusions
Figure 19	Flow Diagram
Figure 20	Plant Plan

Chemical operating costs are estimated as follows:

Sulfuric acid, 66° Be,	17,172 lb/day, $0.016/lb =	$274.75
Caustic soda, 50%	9,855 lb/day, $0.037/lb =	364.64
	Total, $/day =	$639.39

Cost estimates for the equipment as specified in Tables 40 through 44 for this 1.0 MGD size plant, excluding freight, building and land, and assembly and erection, total $428,000.

Cost estimates for erection have been made as follows:

Electrical	$ 9,300
Plumbing	40,000

TABLE 40
HYDROGEN CATION EXCHANGERS
Detailed Specfications, 1.0 MGD

PERFORMANCE:

Total System

Total influent cations, gpg as $CaCO_3$	63.7
Design flow rate, gpm	705
Operating water pressure, psig	75
Number of units	Four

Per Unit

Design flow rate, gpm	176
Peak flow rate, gpm	340
Backwash rate, gpm	313
Cation exchange material, type	Strongly acidic
quantity, cu. ft.	477
capacity, Kgr per cu. ft.	12.5
capacity, Kgr per unit	5,962
Gallons treated per regeneration (includes anion regeneration water)	93,600
Gallons treated to service (net)	83,900
Regenerant, type	H_2SO_4-66° Be
quantity per regeneration	1,431 lbs

SPECIFICATIONS:

Model number	None

Tanks

Tank diameter	120"
Straight side of tank	126"
Design working pressure of tank	100 psi non-code
External surface	Prime painted
Tank lining, material and thickness	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" X 16" Manhole

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Internals

Inlet distributor, design and materials	PVC Header lateral
Regenerant distributor, design and materials	" " "
Underdrain system, design and materials	" " "
Supporting bed	Silica gravel

Piping

Main piping size	6"
Main piping material	Saran lined steel
Main valving arrangement	Nest of diaphragm type
Main valving material	Saran lined cast iron

Table continued on next page

TABLE 40A

HYDROGEN CATION EXCHANGERS continued
Detailed Specifications 1.0 MGD

Control System

Control	Automatic
Initiation of regeneration	Conductance Ratio Meter
Backwash control	Limit stop on valve

Auxiliaries

Meter, size and type	None
Meter register	-
Interconnecting piping between multiple units, inlet & out	6" Saran lined
Pressure gauges	Chemical seal type
Sample cocks	1 pair per unit

Regeneration Equipment

Type of regenerant introduction	Pump
Regenerant introduction strength	2%
Regenerant tank size, bulk storage	10 ft dia X 20 ft
Material of construction	Unlined steel

Electrical Requirements

Volts, Hertz, Phase	115/60/1

ADDITIONAL SPECIFICATIONS:

Conductance Ratio bridges, to determine end of cycle	4 Model RE 18C
Special Sampling manifolds for ratio bridges	Screened header lateral- PVC
Solu-Bridge to monitor acid regenerant strength with selector switch & 4 cells	1 -Model RD 226C
Bypass type ratio of flow meter on each unit	Included
Horizontal concentrated acid storage tank for concrete saddles	Included
Regenerant acid piping system	1/2" Carpenter 20 SS
Waste discharge inter-connecting headers	6" Saran lined
Special Milton Roy acid pump, 3/4 HP T.E. Motor	168 gph

TABLE 41

ANION EXCHANGERS

Detailed Specifications, 1.0 MGD

PERFORMANCE:

Total System

Total influent exchangeable anions, gpg as $CaCO_3$	84.0
Design flow rate, gpm	705
Operating water pressure, psig	50
Number of units	Three

Per Unit

Design flow rate, gpm	235
Peak flow rate, gpm	353
Backwash rate, gpm	314
Anion exchange material, type	Weakly Basic
quantity, cu. ft.	313
capacity, Kgr per cu. ft.	30
capacity, Kgr per unit	9390
Gallons treated per regeneration	111,785
Regenerant, type	NaOH-50% liquid
quantity per regeneration	171.6 gal (1095 lb, 100%)

SPECIFICATIONS:

Model Number	None

Tanks

Tank diameter	120"
Straight side of tank	108"
Design working pressure of tank	100 psi Non-Code
External surface	Prime painted
Tank lining, material and thickness	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" X 16" Manhole

Internals

Inlet distributor, design and materials	PVC header lateral
Regenerant distributor, design and materials	PVC " "
Underdrain system, design and materials	PVC " "
Supporting bed	Silica Gravel

Reproduced from best available copy.

Piping

Main piping size	6"
Main piping material	Saran lined steel
Main valving arrangement	Nest of diaphragm type
Main valving material	Saran lined cast iron

Table continued on next page

TABLE 41A

ANION EXCHANGERS continued

Detailed Specifications, 1.0 MGD

Control System

Control	Automatic
Initiation of regeneration	Conductance Ratio Meter
Backwash control	Limit stop on valve

Auxiliaries

Meter, size and type	None
Meter register	-
Interconnecting piping between multiple units	Included
Pressure gauges (inlet & outlet)	Chemical seal type
Sample cocks	1 pair per unit
Conductivity instrument, type	Conductance Ratio
manufacturer	Beckman
model number	RE-18G

Regeneration Equipment

Type of regenerant introduction	Pump
Regenerant introduction strength	4%
Regenerant tank bulk storage	10' diam X 27'
Material of construction	Unlined Steel

Electrical Requirements

Volts, Hertz, Phase	115/60/1 & 230-460/60/3

ADDITIONAL SPECIFICATIONS:

Special Sampling Manifolds for ratio bridges	Screened header-lateral PVC
Solu-Brdige to monitor caustic regenerant strengths with selector switch & 3 cells	1-Model RD-277C
Bypass rate of flow meter on each unit	Included
Caustic Storage tank for concrete saddles	Unlined
Regenerant caustic piping system	3/4" wrought steel
Special Milton-Roy Caustic pump, 2 HP T.E. Motor	516 gph
Inlet headers for cation effluent	6" Saran lined
Outlet " for anion effluent	6" wrought steel
Waste discharge inter-connecting headers	6" Saran lined

TABLE 42

FORCED DRAFT DEGASIFIERS OR AERATOR

Detailed Specifications, 1.0 MGD

PERFORMANCE:

Total System

Function	Aeration
Influent Iron & Manganese content, ppm	2.4
Influent temperature, °F.	Ambient
Design flow rate, gpm	705
Water pressure at inlet	15 psi
Number of units	One

Per Unit

Design flow rate, gpm	705
Peak flow rate, gpm	705

SPECIFICATIONS:

Reproduced from best available copy.

Tower

Size of tower	96"
Height of tower	144"
Materials of construction	Fir Staves

Internals

Material of packing	Redwood trays
Depth of packing	9 ft
Inlet distributor, design and materials	PVC header-lateral
Support, design and materials	Wood

Auxiliaries

Blower, type , three units needed	Centrifugal
capacity, cfm	3,200 each
static head, inches H_2O	2"
Motor, type	2 HP - ODP
voltage, current, phases	230-460/60/3
Level control, type-for reservoir	Modulating "Leveltrol"
Inlet valve, type	Modulated
material of construction	Cast iron, SS Trim
size	4"

Storage

Concrete reservoir below floor level	Not included
Low water float switch for pump safety cut-off	Included
Forwarding pump, 705 gpm @ 120-100 TDH-30 H.P., ODP Motor, all stainless construction	Included

TABLE 43

PRESSURE FILTERS
Detailed Specifications, 1.0 MGD

PERFORMANCE:

Total System

Design flow rate, gpm	705
Operating water pressure, psig	30 minimum
Number of units	Three
Type of units	Hi-Velocity

Per Unit

Design flow rate, gpm	235
Peak flow rate, gpm	235
Design rate, gpm/ft^2 of filter area	10
Backwash rate, gpm	350
Filter media, type	Hi-Velocity
quantity, cu. ft.	65
depth of bed, inches	33

SPECIFICATIONS:

Model Number	None

Tanks

Tank diameter	66"
Straight side of tank	84"
Design working pressure of tank	100 psi Non-Code
External surface	Prime Painted
Internal surface	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" X 16" Manhole

Internals

Inlet distributor, design and materials	PVC Header-lateral
Underdrain system, design and materials	PVC " "
Supporting bed	Gravel

Piping

Main piping size	4" X 6"
Main piping material	Saran Lined
Main valving arrangement	Nest Of Diaphragm Valves

Control System

Control	Automatic
Initiation of regeneration	Timeclock
Backwash control	Limit stop on valves

Table continued on next page

Reproduced from best available copy.

TABLE 43A

PRESSURE FILTERS (Cont'd)

Detailed Specifications, 1.0 MGD

Auxiliaries	
Interconnecting piping between multiple units	6" wrought steel
Pressure gauges	3 pair included
Sample cocks	3 pair included
Electrical Requirements	
Voltz	115
Hertz	60
Phase	1

ADDITIONAL SPECIFICATIONS:

Bypass type rate of flow meter two on each unit	Included
Waste discharge inter-connecting headers	6" Saran lined

TABLE 44

MISCELLANEOUS ITEMS INCLUDED:

Detailed Specifications, 1.0 MGD

Air Compressor for all plant control needs
Modulating pH Meter, alarm, pH flow chamber and electrodes with milliamp output to control pH adjustment pump
Modulating pH Meter as above but with tranducer added for 3-15 psi air signal to control final acid blending - neutralizing valve
Two-pen strip chart recorder for pH instruments
Alloy 20 blending valve for cation effluent with controls to modulate feed from 2-15 psi air signal
Caustic feed pump to adjust pH prior to aeration, equipped with controller station to modulate feed from milliamp input
All inter-plant piping needed to inter-connect ion exchange units, aerator & filters, Saran lined, black wrought steel, & PVC, 8", 6", 1", & 3/4"

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The following items are specifically excluded:

1. The building, its foundations, concrete reservoir below aerator
2. Auxilliary plumbing and plumbing fixtures for the building
3. Electrical wiring of building & connections between electrical controls: **costs separately estimated**
4. All pump starters: **costs separately estimated**

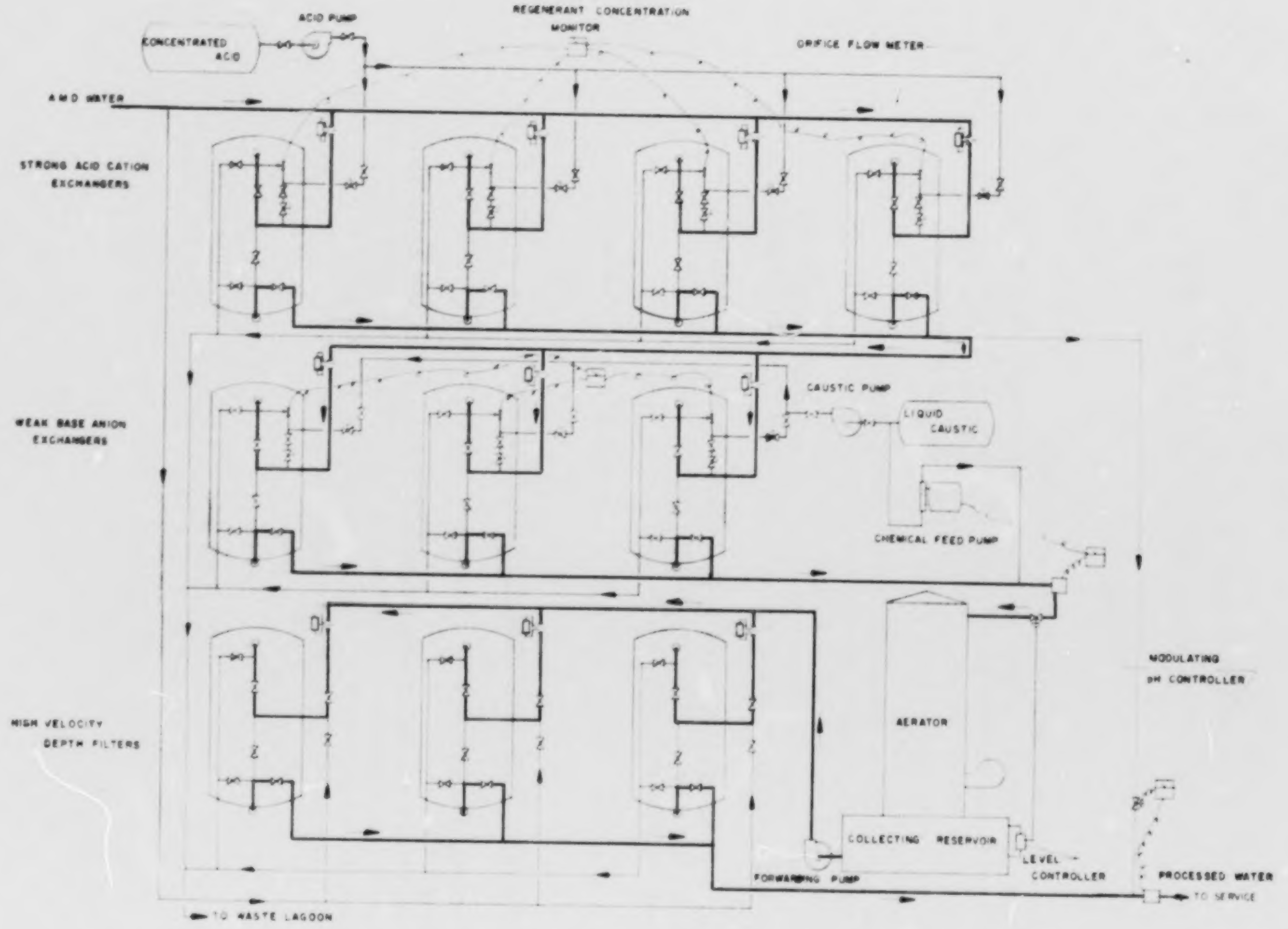
CONCENTRATED ACID
ACID PUMP
REGENERANT CONCENTRATION MONITOR
ORIFICE FLOW METER
AMD WATER
STRONG ACID CATION EXCHANGERS
CAUSTIC PUMP
LIQUID CAUSTIC
WEAK BASE ANION EXCHANGERS
CHEMICAL FEED PUMP
MODULATING pH CONTROLLER
AERATOR
HIGH VELOCITY DEPTH FILTERS
COLLECTING RESERVOIR
LEVEL CONTROLLER
FORWARDING PUMP
PROCESSED WATER
TO SERVICE
TO WASTE LAGOON

FIGURE 19. AMD Treatment Plant Flow Diagram, 2-Resin System, 1.0 MGD

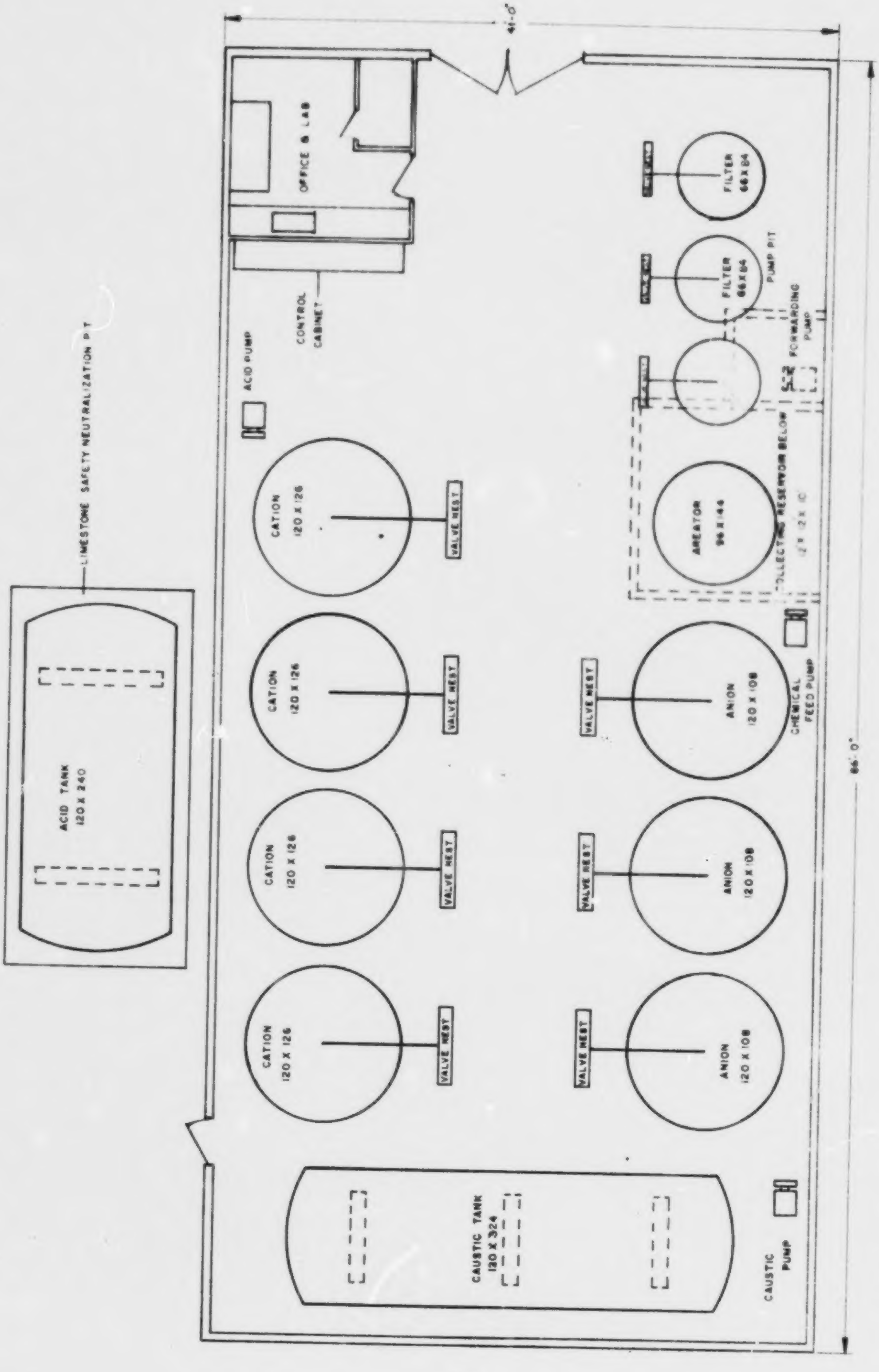
LIMESTONE SAFETY NEUTRALIZATION PT
ACID TANK 120 X 240
41'-0"
OFFICE & LAB
CONTROL CABINET
ACID PUMP
CATION 120 X 126
CATION 120 X 126
CATION 120 X 126
CATION 120 X 126
VALVE NEST
VALVE NEST
VALVE NEST
VALVE NEST
VALVE NEST
VALVE NEST
VALVE NEST
FILTER 66 X 84
FILTER 66 X 84
PUMP PIT
FORWARDING PUMP
AREATOR 96 X 144
COLLECTING RESERVOIR BELOW 12' X 12' X 10'
ANION 120 X 108
ANION 120 X 108
ANION 120 X 108
CHEMICAL FEED PUMP
86'-0"
CAUSTIC TANK 120 X 324
CAUSTIC PUMP

FIGURE 20. AMD Treatment Plant Plan, 2-Resin System, 1.0 MGD

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SECTION 17

TREATMENT PLANT DESIGN - MODIFIED DESAL PROCESS

This AMD treatment system is an automatic ion exchange-precipitation-filtration process to produce potable water from acid mine drainage containing excessive amounts of iron and sulfuric acid. The method used is a modification of the Desal process and is essentially one of anion exchange followed by precipitation and filtration with final pH neutralization.

Section 11 of this report discussed the preliminary tests on only the bicarbonate form of the weak base resin. Section 15 discussed the use of this resin and post treatment to effect a complete treatment of AMD to produce potable water.

The design parameters for plants employing this process were presented in Table 29. Those parameters have been used to design plants for the treatment of acid mine drainage water. Plants of three sizes have been designed; namely, 0.1, 0.5, 1.0 MGD (million gallons per day).

Summary of costs:

Figure 21 is a plot of the equipment costs, in hundreds of thousands of dollars for the various size plants to treat AMD by this modified Desal system. These prices are for the equipment listed in the detailed specifications, but exclude freight, building and land, assembly and erection.

Figure 22 is a plot of the estimated chemical operating costs, in hundreds of dollars per day for the plants to treat the AMD by the modified Desal system. All utility costs are excluded.

Figures 23 and 24 plot, respectively, estimated erection labor costs for the electrical and plumbing requirements.

General Discussion of Plants:

The equipment will consist essentially of three pressure vessels (tanks) containing anion exchange resin, an aerator, a reactor-clarifier, and two or more tanks containing granular minerals for final filtration of the product water.

The operation of this treatment plant is predicated on the assumption that the AMD will be supplied to the system at 75-100 psi pressure, free of turbidity, and will conform to the raw AMD analysis given elsewhere in this report.

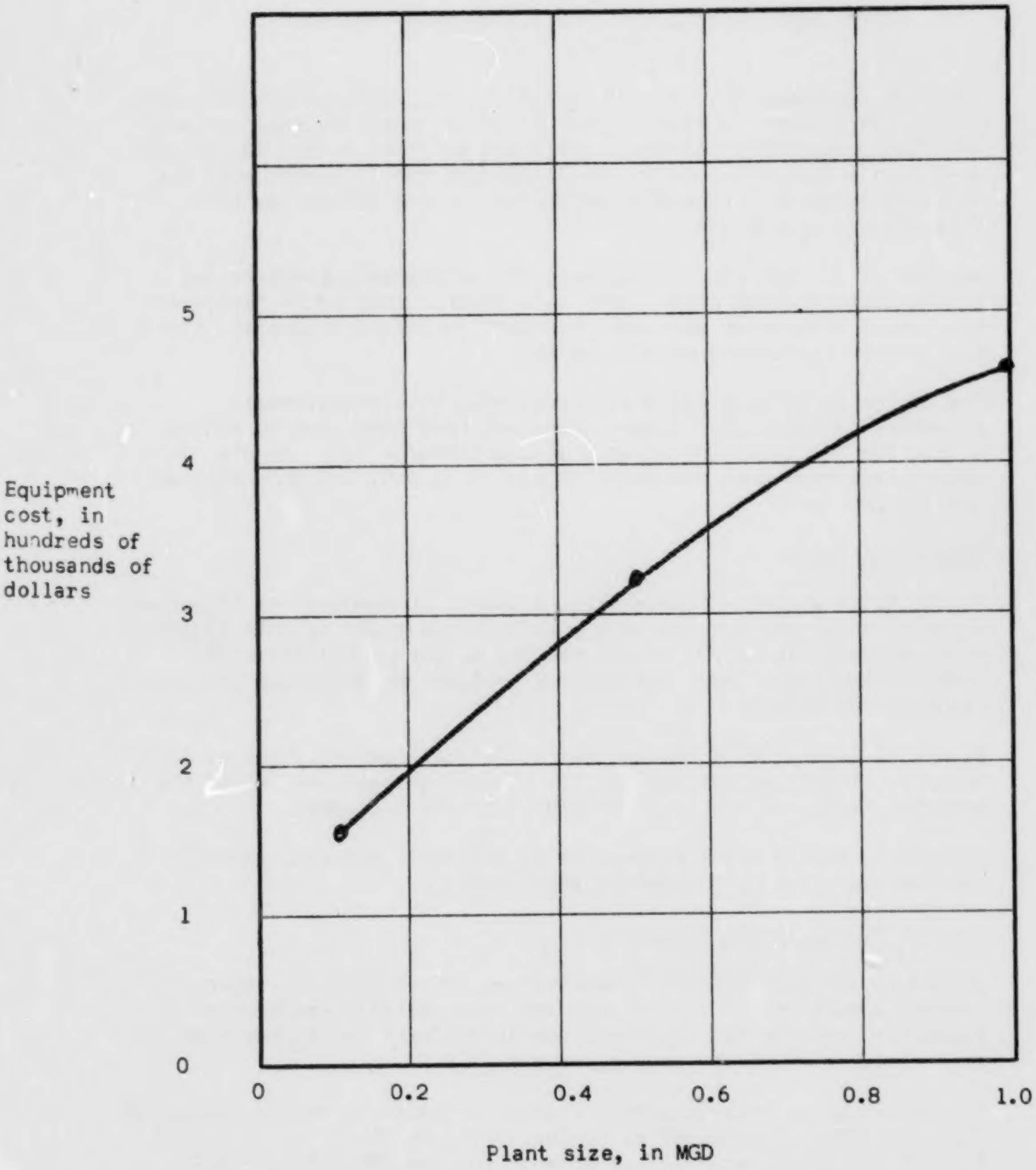

Figure 21. Cost estimates for unassembled, unerected equipment to treat AMD by the modified Desal system.

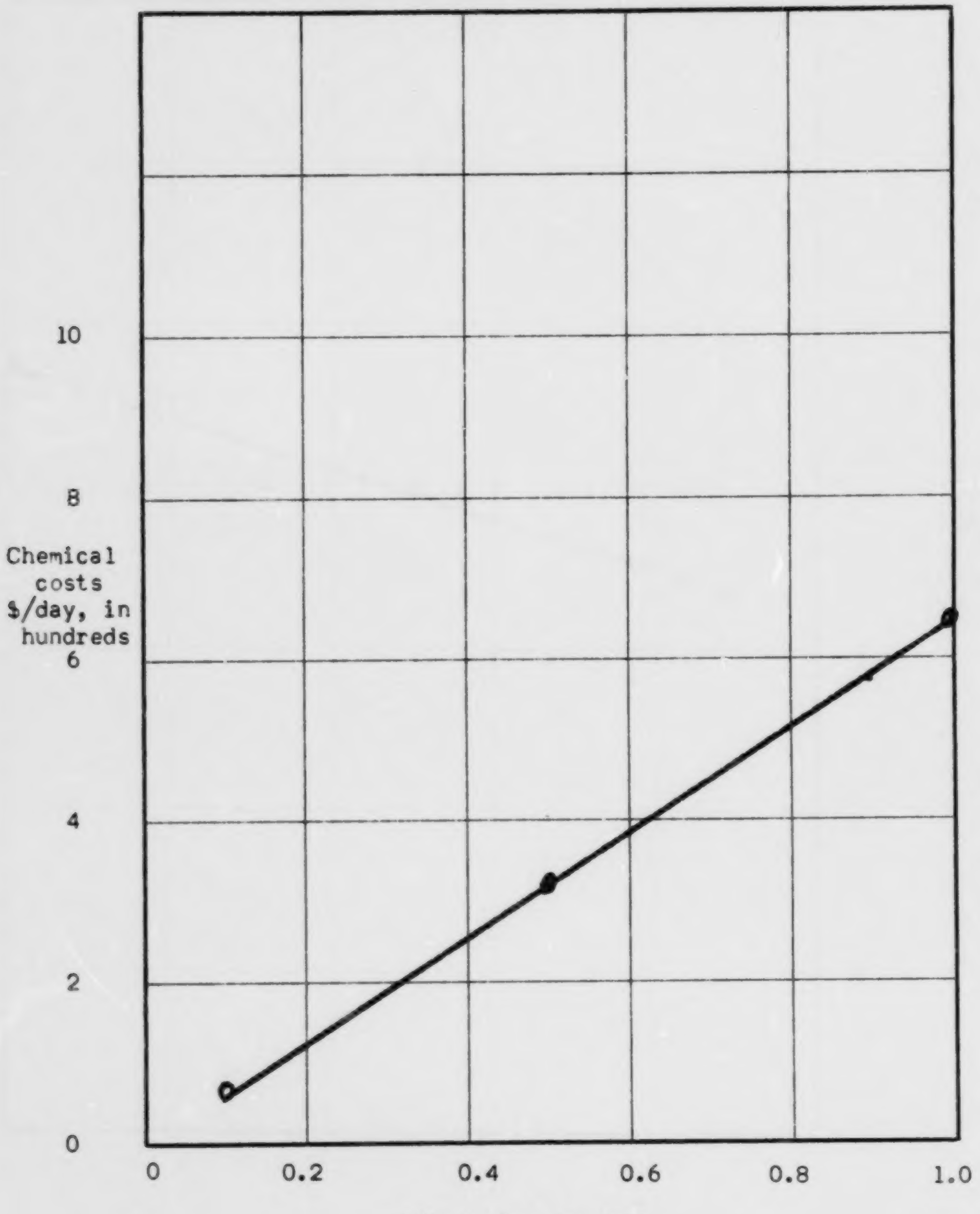

Figure 22 Estimates of daily chemical operating costs to treat AMD by the modified Desal system.

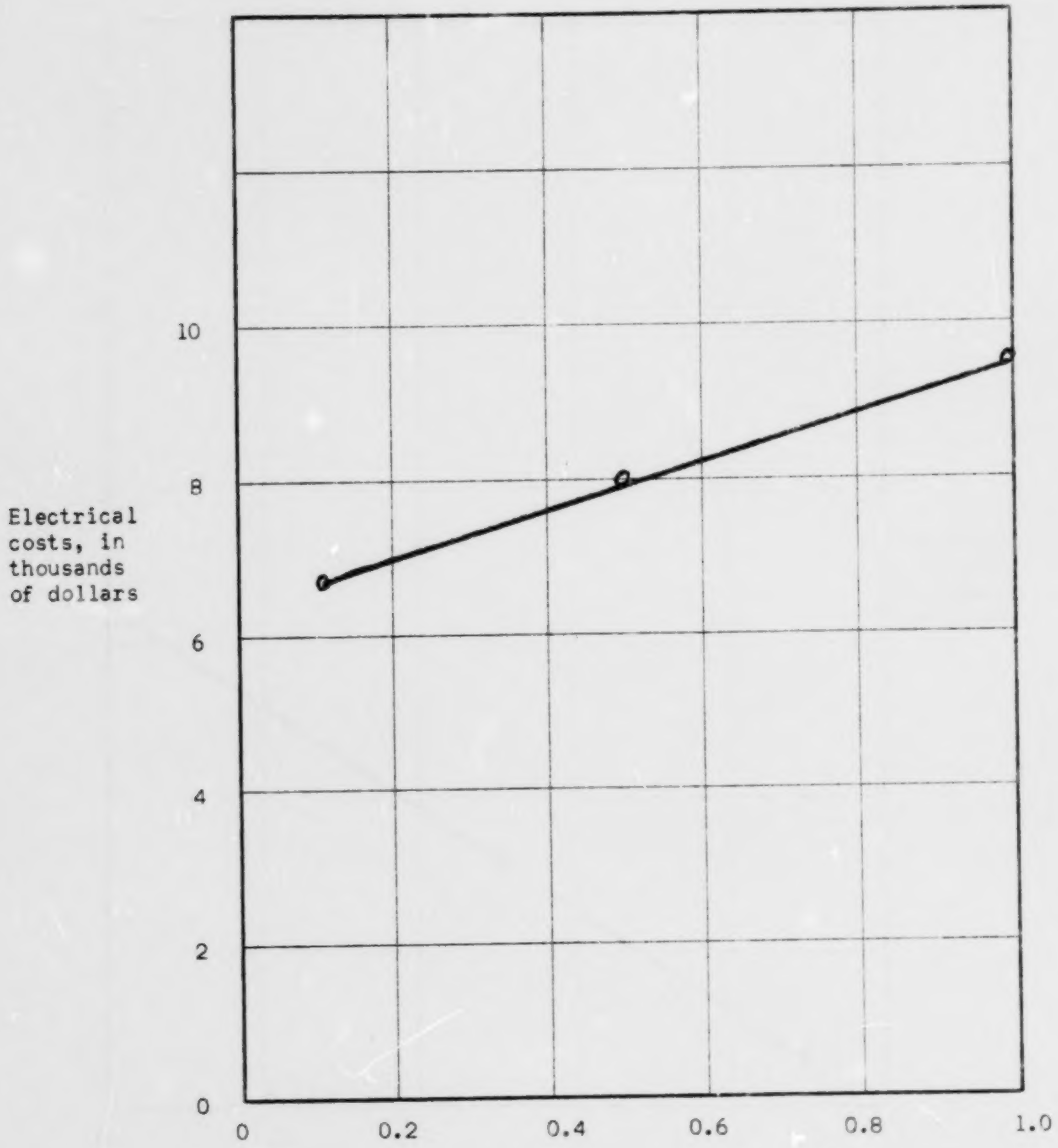

Figure 23 Estimates of electrical labor costs for erection of plants to treat AMD by the modified Desal system.

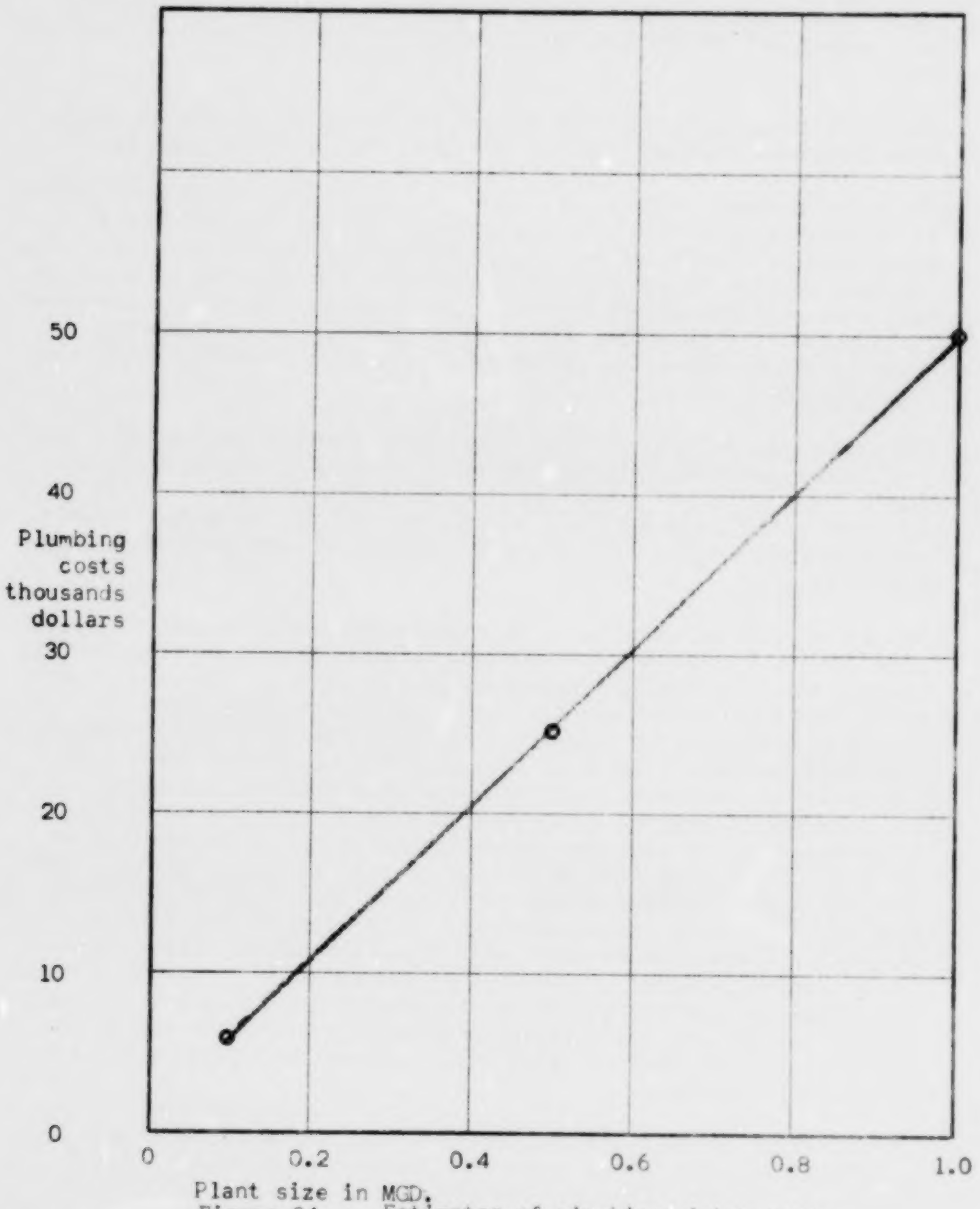

Figure 24 Estimates of plumbing labor costs for assembly and erection of plants to treat AMD by the modified Desal system.

The exchanger vessels will be arranged for parallel flow with a common inlet header and a common outlet header. The anion exchange resin will remove all of the acid and much of the iron and sulfate from the AMD.

After the anion exchange, the metal ions are associated essentially with bicarbonate anions, making the water suitable for coagulation.

The anion exchange vessels are operated upflow in service at 2 gpm/cu ft to prevent excessive pressure loss due to iron precipitation. Upper take-off manifolds must be screened to prevent resin loss. The effluent quality is monitored by pH meters, the pH being recorded for each exchanger throughout the cycle. The exhaustion of the exchanger is signaled by a pH drop. The signal actuates the regeneration controls which remove the exchanger from service, thereby initiating regeneration.

The first step of regeneration is a normal backwash through a separate collecting manifold. The backwash and subsequent regeneration will result in a waste which must be treated before disposal. While it is not a part of this study or plant design, all of the backwash wastes and regenerant wastes in rinse water should be collected in a common waste lagoon.

The anion exchange resin will be regenerated downflow with 4% caustic soda (sodium hydroxide). A caustic-resistant pump will transfer 50% caustic solution from the bulk storage facilities. The caustic will be diluted to the desired 4% strength with water taken from the clarifier effluent. A separate forwarding pump is used to transfer this dilution water, which is also used for regenerant rinse. At completion of the rinse, the anion exchange resin is converted to the bicarbonate form by recirculating a carbon dioxide solution. Rinse water in the exchanger is recirculated by a separate pump at 100 psi. The closed loop recirculation rate is 0.5 in gpm per cubic foot of resin. Carbon dioxide is injected into this recirculating flow until complete resin conversion to the bicarbonate form is achieved. A break in pH tells when this is reached. High pressure recirculation is used to provide a strong solution of carbon dioxide. The estimated requirement for carbon dioxide is 3.3 pounds per cubic foot of resin. The fully regenerated anion exchange vessel is now automatically returned to service, or standby, as dictated by individual plant design.

The AMD which has been treated by the anion exchange resin next enters the aerator, where oxygen is absorbed from the air. This oxygen will oxidize the iron and manganese. Aeration is by means of a standard tray type forced draft counter-flow aerator. Water exists by gravity from the aerator into the reactor-clarifier.

The reactor-clarifier tank will be of concrete construction, mostly below ground level. The clarifier should project above the ground by 18-24" to reduce personnel hazards while permitting gravity flow. Lime (slaked or unslaked calcium oxide depending upon plant size) and coagulant aid are used to remove all of the iron and manganese, while reducing calcium, magnesium, and alkalinity to low levels. Resulting precipitates accumulate in a sludge blanket that is maintained in the reactor-clarifier, which is designed with a rise rate of 1.0 gpm/cu ft. The effluent pH from the clarifier is 10.1. This pH must be reduced for final use; we have designed this to be done after filtration.

The clarifier outflow proceeds by gravity to the clear well which is at a lower level than the aerator. A float control in the clear well controls the aerator inlet valve, thus controlling the operation of the aerator, clarifier and clear well. The bottom of the aerator is higher than the top of the clarifier and the following clear well, so flow proceeds through these components by gravity.

Final product filtration requires a pump to transfer water from the clear well of the reactor-clarifier to the filters. The filters remove any insoluble particles that have carried over from the clear well. Acid is fed to the filtered product by a chemical pump controlled by a pH meter and recorder. Thus, a final product is delivered which meets the mineral requirements for potable water.

The filters will be backwashed and rinsed with treated water from the aerator clear well (as in service). The backwash and rinse effluents will be sent to the waste lagoon. The backwash and rinse operations will be time controlled, automatically initiated once a day for each filter. The filters are so sized that removal of one filter from service will not overload the system or interrupt the flow of product water.

Designs and specifications for each plant (0.1, 0.5 and 1.0 MGD) are detailed on the following pages.

Details of Plant Designs:

Plant to produce 0.1 MGD:

The equipment will consist essentially of three anion exchange vessels, one aerator, one reactor-clarifier and clear well, two final filters.

Three anion exchange vessels are required although the plant is designed to schedule only one vessel in service at a time. Three are required because the total regeneration time of one vessel is longer than its service run. The complete cycle for one vessel is three hours service, four hours regeneration, two hours standby. Thus, if only one vessel is on stream at a time and takes the full plant flow, eight regenerations per day is a workable schedule and produces the required plant output.

The detailed specifications of the equipment for this plant to treat 0.1 MGD are on the following pages.

Table 45	Anion Exchangers
Table 46	Forced Draft Degasifiers or Aerator
Table 47	Reactor-Clarifier
Table 48	Pressure Filters
Table 49	Miscellaneous Items and Exclusions
Figure 25	Flow Diagram
Figure 26	Plant Plan

Chemical operating costs are estimated as follows:

Caustic Soda, 50%, 1,200 lb/day, $0.037 per pound	= $44.40
Carbon Dioxide, 992 lb/day, $0.015 per pound	= 14.88
Lime, 350 lb/day, $0.01 per pound	= 3.50
Total, $/day	= $62.78

Cost estimates for the equipment as specified in Tables 45 through 49 for this 0.1 MGD size plant, excluding freight, building and land, and assembly and erection, total $156,000.

Cost estimates for erection have been made as follows:

Electrical	$ 9,300
Plumbing	5,000

TABLE 45

ANION EXCHANGERS

DETAILED SPECIFICATIONS, 0.1 MGD

PERFORMANCE:

Total System	
Total influent exchangeable anions, gpg as $CaCO_3$	84.0
Design flow rate, gpm	70
Operating water pressure, psig	40, minimum
Number of units	Three
Per Unit	
Design flow rate, gpm	74.5
Peak flow rate, gpm	74.5
Backwash rate, gpm	19
Anion exchange material, type	IRA-68
quantity, cu ft	37.5
capacity, Kgr per cu ft	30
capacity, Kgr per unit	1,125
Gallons treated per regeneration	13,390
Gallons net to service	12,610
Regenerant quantity per regeneration, lbs.	
Sodium Hydroxide	150
Carbon Dioxide	124

SPECIFICATIONS:

Model Number	None
Tanks	
Tank diameter	42"
Straight side of tank	108"
Design working pressure of tank	100 psi Non-Code
External surface	Prime painted
Tank lining, material and thickness	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" x 16" manhole
Internals	
Backwash collecting manifold, design & materials	PVC Header lateral
Treated water collecting manifold, design & materials	PVC-Screened header lateral
Regenerant distributor, design and materials	PVC header lateral
Underdrain system, design and materials	PVC header lateral
Supporting bed	Silica gravel

TABLE 45A ANION EXCHANGER (continued)

Piping

Main piping size	2½" x 1½"
Main piping material	Saran lined steel
Main valving arrangement	Nest of auto diaph type
Main valving material	Saran lined cast iron

Control System

Control	Automatic
Initiation of regeneration	pH meter
Backwash control	Limit stop on valves

Auxiliaries

Meter, size and type	None
Meter register	--
Interconnecting piping between multiple units, inlet and outlet	3" saran lined
Pressure gauges	chemical seal type
Sample cocks	1 pair per unit
Conductivity instrument, type	None
manufacturer	--
model number	--

Regeneration Equipment

Type of regenerant introduction	Positive displace pump
Regenerant introduction strength	4%
Regenerant tank size, vertical bulk storage	8' dia x 7'
Material of construction	Unlined steel

Electrical Requirements

Volts, Hertz, Phase	115/60/1 & 230-460/60/3

ADDITIONAL SPECIFICATIONS:

Regenerant caustic pump, Milton Rog 1/2 HP	60 gph
Regenerant caustic piping system	1" wrought steel
Regeneration water header from clear well	1¼" wrough steel
Regeneration water pump - Gould #3196 all iron centrifugal for 18 gpm @ 50' head with 1.5 HP ODP motor	Included
Waste discharge piping	1½" saran lined steel
Bypass type rate of flow meters, 2 for each tank	1¼"
Bypass type rate of flow meter, 1 for product output	2½"
Beckman Model 940 pH meter with flow chamber, and electrodes to detect endpoints	1 for each tank
Three pen strip chart recorder	4"
6 ton liquid CO_2 storage container with 4.5 KW vaporizer and regulator	Included

TABLE 45B ANION EXCHANGERS (continued)

1½" stainless steel sparger to dispurse CO_2 gas in recirculating line	1 per tank
Pressure relief and suction relief on each tank	Included
Piping from CO_2 storage to points of use, wrought steel	Included
Pump to recycle water during CO_2 saturation, Gould #3196 all stainless steel, 15 HP, 3600 rpm, 230-460/60/3 ODP motor for 19 gpm @ 100 psi	1 for each tank

TABLE 46

FORCED DRAFT DEGASIFIERS OR AERATOR
DETAILED SPECIFICATIONS, 0.1 MGD

PERFORMANCE:

Total System

Function	Aeration
Influent iron and manganese content, ppm maximum	140
Influent temperature, oF	Ambient
Design flow rate, gpm	70
Water pressure at inlet	15 psi
Number of units	One

Per Unit

Design flow rate, gpm	75
Peak flow rate, gpm	75

SPECIFICATIONS:

Tower

Size of tower	36" diameter
Height of tower	144"
Materials of construction	Fir Staves

Internals

Material of packing	Red wood tray
Depth of packing	9 feet
Inlet distributor, design and materials	PVC header lateral
Support, design and materials	Wood

Auxiliaries

Blower, type	Centrifugal
capacity, cfm	840
static head, inches H_2O	2
Motor, type	1 HP - ODP
voltage, current, phases	230-460/60/3
Level control, type for clear well	Modulating "levetrol"
Inlet valve, type	Modulated
material of construction	Cast iron - SS trim
size	1.5"

STORAGE: None. Water flows directly to Reactor-Clarifier and from there to concrete clear well. Aerator is located above clear well. Water level in clear well controls aerator inlet valve and thus also feed to reactor-clarifier. Modulating level control is itemized above.

TABLE 47

REACTOR-CLARIFIER
DETAILED SPECIFICATION, 0.1 MGD

Size of Tank, diameter x side wall depth	12' x 12'
Tank construction	Concrete
Design flow rate, gpm	75
Rise rate, gpm per sq. ft.	1
Detention period, minutes	130
Bottom slope to central sludge cone	1 in 12
Provisions for mixing, internal recirculation, flocculation, settling, clarification, positive sludge thickening and removal	Included
Lime feeder and coagulant feeder	Included
Sludge rake drive mechanism fully enclosed with motor	0.5 HP
Turbine driving mechanism with motor	0.5 HP
Beam type superstructure spanning tank	36" wide floor plate
Conical reaction chamber	Included
Peripheral effluent collection launder	Included
Sampling pipes	Included
Plug type sludge valve with pneumatic cylinder operator	Included
Piping from aerator to reactor-clarifier, welded	3" - Included
Piping from reactor-clarifier to clear well, welded	3" - Included
Estimated lime dosage, lbs. of 100% CaO per 1000 gal.	3.2
lbs. of 100% $Ca(OH)_2$ per day	458

Clarifier will be mostly below floor, only the top 18" to 24" of the concrete tank will be above floor level.

CLEAR WATER

Concrete construction, below floor level of equipment building, size 6' x 5' x 10' deep. Nominal capacity 1,800 gallons. Concrete construction is not included as part of equipment design cost and should be included under building construction. Two pumps are located in the pump pit adjacent to clear well. One is for regenerating the anion units and is described under Anion Exchangers. The other is a transfer or forwarding pump for regular service operation. It is a Gould #3196 all iron centrifugal pump for 70 gpm at 50 psi with 5 HP, 230-460/60/3 1750 rpm ODP motor.

TABLE 48

PRESSURE FILTERS
DETAILED SPECIFICATIONS, 0.1 MGD

PERFORMANCE:

Total System

Design flow rate, gpm	70 gpm
Operating water pressure, psig	30 psi min.
Number of units	Two
Type of units	Hi-Velocity

Per Unit

Design flow rate, gpm	35
Peak flow rate, gpm	70
Design rate, gpm/ft^2 of filter area	7.2
Backwash rate, gpm	73
Filter media, type	Hi-Velocity
quantity, cu ft	13.5
depth of bed, inches	33

SPECIFICATIONS:

Model Number	None

Tanks

Tank diameter	30"
Straight side of tank	60"
Design working pressure of tank	100 psi Non-Code
External surface	Prime painted
Internal surface	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" x 16" manhole

Internals

Inlet distributor, design and materials	PVC Header lateral
Underdrain system, design and materials	PVC Header lateral
Supporting bed	Gravel

Piping

Main piping size	2"
Main piping material	PVC
Main valving arrangement	Nest of Auto. valves

TABLE 48A PRESSURE FILTERS (continued)

Control System

Control	Automatic
Initiation of regeneration	Time Clock
Backwash control	Limit stop on valves

Auxiliaries

Interconnecting piping between multiple units	2" wrought steel
Pressure gauges	3 pr included
Sample cocks	3 pr

Electrical Requirements

Volts, Hertz, Phase	115/60/1

ADDITIONAL SPECIFICATIONS:

Bypass type rate of flow meters, 2 on each unit	Included
Waste discharge interconnecting headers	3" PVC
Filters are backwashed with raw AMD water. Valve nest, waste lines, and AMD supplies are corrosion resistant for the service	Included

TABLE 49

MISCELLANEOUS ITEMS
DETAILED SPECIFICATIONS, 0.1 MGD

Air compressor for all plant control needs	Included
Modulating pH meter, alarm, pH flow cell and electrodes with milliamps output to control pH correction pump. Also strip chart recorder	Included
Acid feed pump with variable speed drive modulated by milliamp signal from pH meter	Included
Acid storage tank will be container in which it is shipped	Not included
Acid supply and feed piping from storage tank to pump and point of feed	½" carpentier 20SS
All interplant piping needed to interconnect ion exchange units, aerator, clarifier and filters in saran lined and black wrought steel as required	Included

* *

The following items are specifically excluded.

1. The building, its foundations, concrete reservoir, concrete clarifier tank, and concrete tank saddles.
2. Auxiliary plumbing and plumbing fixtures for the building.
3. Electrical wiring of building and connections between electrical controls.
4. All pump starters.
5. Installation and erection of equipment.
6. The equipment to handle bulk unslaked lime is not considered to be included as part of this plant design.

NOTE: The handling and supply of the large quantities of carbon dioxide required would well be the subject of a separate study. The reader should realize that other methods of supply might be usable and should conduct his own evaluation.

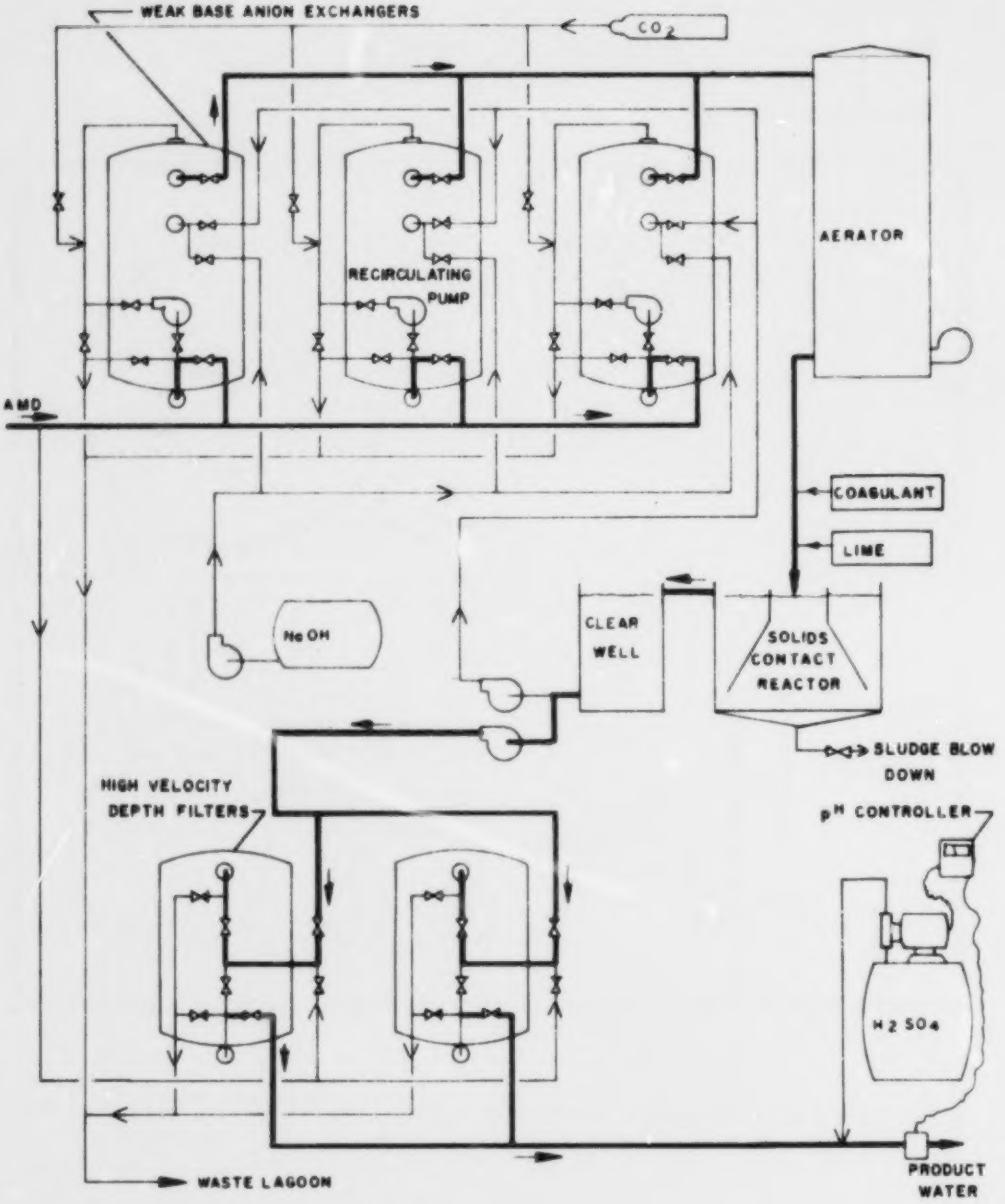
WEAK BASE ANION EXCHANGERS
CO2
AERATOR
RECIRCULATING PUMP
AMD
COAGULANT
LIME
NaOH
CLEAR WELL
SOLIDS CONTACT REACTOR
SLUDGE BLOW DOWN
HIGH VELOCITY DEPTH FILTERS
pH CONTROLLER
H2SO4
PRODUCT WATER
WASTE LAGOON

FIGURE 25. AMD Treatment Plant Flow Diagram, Modified Desal System, 0.1 MGD

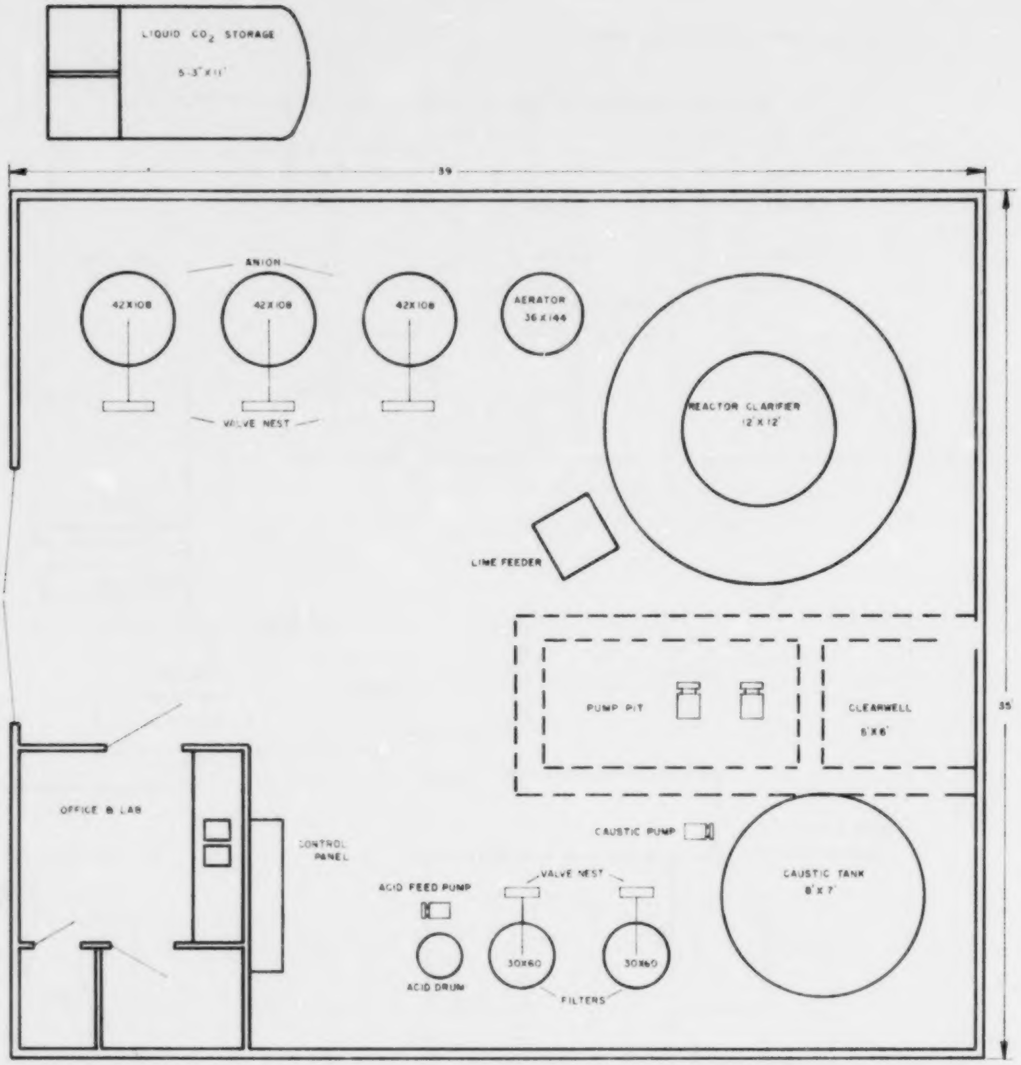
LIQUID CO_2 STORAGE
5-3" X 11'
39
35'
ANION
42X108
42X108
42X108
VALVE NEST
AERATOR
36X144
REACTOR CLARIFIER
12' X 12'
LIME FEEDER
PUMP PIT
CLEARWELL
8'X8'
OFFICE & LAB
CONTROL
PANEL
CAUSTIC PUMP
CAUSTIC TANK
8' X 7'
VALVE NEST
ACID FEED PUMP
30X60
30X60
ACID DRUM
FILTERS

FIGURE 26. AMD Treatment Plant Plan, Modified Desal System, 0.1 MGD

Plant to produce 0.5 MGD:

The equipment will consist essentially of three anion exchange vessels, one aerator, one reactor-clarifier and three final filters.

Three anion exchange vessels are incorporated in the design although only one will be in service operation at a time. Three are required because total regeneration time of one vessel is longer than its service run. The complete cycle for one tank is three hours service, four hours regeneration, two hours standby. Thus, if only one tank is on stream at a time and takes the full plant flow, eight regenerations per day is a workable schedule and produces the required plant output of treated water.

The detailed specifications of the equipment for this plant to treat 0.5 MGD are on the following pages.

Table 50	Anion Exchangers
Table 51	Forced Draft Degasifier or Aerator
Table 52	Reactor-Clarifier
Table 53	Pressure Filters
Table 54	Miscellaneous Items and Exclusions
Figure 27	Flow Diagram
Figure 28	Plant Plan

Chemical operating costs are estimated as follows:

Caustic Soda, 50%, 5,984 lb/day, $0.037 per pound	=	$221.41
Carbon Dioxide, 4,936 lb/day, $0.015 per pound	=	74.04
Lime, 1,450 lb/day, $0.01 per pound	=	14.50
Tota, $/day	=	$309.95

Cost estimates for the equipment as specified in Tables 50 through 54 for this 0.5 MGD size plant, excluding freight, building and land, and assembly and erection, total $323,000.

Cost estimates for erection have been made as follows:

Electrical	$ 7,800
Plumbing	25,000

TABLE 50
ANION EXCHANGERS

DETAILED SPECIFICATIONS, 0.5 MGD

PERFORMANCE:

Total System

Total influent exchangeable anions, gpg as $CaCO_3$	84.0
Design flow rate, gpm	350
Operating water pressure, psig	40, minimum
Number of units	Three

Per Unit

Design flow rate, gpm	371
Peak flow rate, gpm	371
Backwash rate, gpm	100
Anion exchange material, type	IRA-68
quantity, cu ft	187
capacity, Kgr per cu ft	30
capacity, Kgr per unit	5,610
Gallons treated per regeneration	67,785
Gallons net to service	503,100
Regenerant quantity per regeneration, lbs	
Sodium Hydroxide	748
Carbon Dioxide	617

SPECIFICATIONS:

Model Number	None

Tanks

Tank diameter	96"
Straight side of tank	108"
Design working pressure of tank	100 psi Non - Code
External surface	Prime painted
Tank lining, material and thickness	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" x 16" manhole

Internals

Backwash collecting manifold, design & materials	PVC header lateral
Treated water collecting manifold, design & materials	PVC-Screened header lateral
Regenerant distributor, design and materials	PVC header lateral
Underdrain system, design and materials	PVC header lateral
Supporting bed	Silica gravel

TABLE 50A ANION EXCHANGERS (Continued)

Piping	
Main piping size	3" x 2"
Main piping material	Saran lined steel
Main valving arrangement	Nest of auto diaph type
Main valving material	Saran lined cast iron
Control System	
Control	Automatic
Initiation of regeneration	pH meter
Backwash control	Limit stop on valves
Auxiliaries	
Meter, size and type	None
Meter register	--
Interconnecting piping between multiple units, inlet and outlet	3" Saran lined
Pressure gauges	Chemical seal type
Sample cocks	1 pair per unit
Conductivity instrument, type	None
manufacturer	--
model number	--
Regeneration Equipment	
Type of regenerant introduction	Positive displace. pump
Regenerant introduction strength	4%
Regenerant tank size, horizontal bulk storage	10' dia x 18'
Material of construction	Unlined Steel
Electrical Requirements	
Volts, Hertz, Phase	115/60/1 & 230-460/60/3
ADDITIONAL SPECIFICATIONS:	
Regenerant caustic pump, Milton Roy, 2HP	300 gph
Regenerant caustic piping system	1" wrought steel
Regeneration water header from clear well	2½" wrough steel
Regeneration water pump-Gould #3196 all iron centrifugal for 90 gpm @ 50' head with 3 HP ODP motor	Included
Waste discharge piping	3" Saran lined steel
Bypass type rate of flow meters, 2 for each tank	3"
Bypass type rate of flow meter, 1 for product output	6"
Beckman Model 940 pH meter with flow chamber, and electrodes to detect endpoints	1 for each tank
Three pen strip chart recorder	3"
12 ton liquid CO_2 storage container with 2 HP compressor, 9 KW vaporizer and regulator	Included

Table continued on next page

TABLE 50 B ANION EXCHANGERS (continued)

3" stainless steel sparger to dispurse CO_2 gas in recirculating line	1 per tank
Pressure relief and suction relief on each tank	Included
Piping from CO_2 storage to points of use, wrought steel	Included
Pump to recycle water during CO_2 saturation, Gould #3196 all stainless steel, 5 HP, 3600 rpm, 230-460/60/3 ODP motor for 95 gpm @ 100 psi	1 for each tank

TABLE 51

FORCED DRAFT DEGASIFIERS OR AERATOR DETAILED SPECIFICATIONS, 0.5 MGD

PERFORMANCE:

Total System

Function	Aeration
Influent iron and manganese content, ppm maximum	140
Influent temperature, °F	Ambient
Design flow rate, gpm	350
Water pressure at inlet	15 psi
Number of units	One

Per Unit

Design flow rate, gpm	738
Peak flow rate, gpm	638

SPECIFICATIONS:

Tower

Size of tower	72" diameter
Height of tower	144"
Materials of construction	Fir Staves

Internals

Material of packing	Red wood tray
Depth of packing	9 feet
Inlet distributor, design and materials	PVC header lateral
Support, design and materials	Wood

Auxiliaries

Blower, type two units needed	Centrifugal
capacity, cfm	2100 each
static head, inches H_2O	2
Motor, type	1.5 HP - O.D.P.
voltage, current, phases	230-460/60/3
Level control, type for clear well	Modulating "levetrol"
Inlet valve, type	Modulated
material of construction	Cast iron - SS trim
size	3"

STORAGE: None. Water flows directly to Reactor-Clarifier and from there to concrete clear well. Aerator is located above clear well. Water level in clear well controls aerator inlet valve and thus also feed to reactor-clarifier. Modulating level control is itemized above.

TABLE 51
REACTOR-CLARIFIER
DETAILED SPECIFICATION, 0.5 MGD

Size of Tank, diameter x side wall depth	24' x 12'
Tank construction	Concrete
Design flow rate, gpm	371
Rise rate, gpm per sq ft	1
Detention periods, minutes	105
Bottom slope to central sludge cone	1 in 12
Provisions for mixing, internal recirculation, flocculation, settling, clarification, positive sludge thickening and removal	Included
Lime slaker and coagulant feeder	Included
Sludge rake drive mechanism fully enclosed with motor	1.0 HP
Turbine driving mechanism with motor	0.75 HP
Beam type superstructure spanning tank	36" wide floor plate
Conical reaction chamber	Included
Peripheral effluent collection launder	Included
Sampling pipes	Included
Plug type sludge valve with pneumatic cylinder operator	Included
Piping from aerator to reactor-clarifier, welded	6" - Included
Piping from reactor-clarifier to clear well, welded	6" - Included
Estimated lime dosage, lbs. of 100% CaO per 1000 gal.	3.2
lbs. of 100% CaO per day	1,710

Clarifier will be mostly below ground, only the top 18" to 24" of the concrete tank will be above ground level

CLEAR WATER

Concrete construction, below floor level of equipment building, size 7.5' x 18' x 10' deep. Nominal capacity 8,000 gallons. Concrete construction is not included as part of equipment design cost and should be included under building construction. Two pumps are located in the pump pit adjacent to clear well. One is for regenerating the anion units and is described under Anion Exchangers. The other is a transfer or forwarding pump for regular service operation. It is a Gould #3196 all iron centrifugal pump for 351 gpm at 50 psi with 20 HP, 230-460/60/3 1750 rpm ODP motor.

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TABLE 53

PRESSURE FILTERS
DETAILED SPECIFICATIONS, 0.5 MGD

PERFORMANCE:

Total System

Design flow rate, gpm	350 gpm
Operating water pressure, psig	30 psi min.
Number of units	Three
Type of units	Hi-Velocity

Per Unit

Design flow rate, gpm	117
Peak flow rate, gpm	176
Design rate, gpm/ft^2 of filter area	9.4
Backwash rate, gpm	185
Filter media, type	Hi-Velocity
quantity, cu ft	34.5
depth of bed, inches	33

SPECIFICATIONS:

Model Number	None

Tanks

Tank diameter	48"
Straight side of tank	72"
Design working pressure of tank	100 psi Non-Code
External surface	Prime painted
Internal surface	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" x 16" manhole

Internals

Inlet distributor, design and materials	PVC Header lateral
Underdrain system, design and materials	PVC Header lateral
Supporting bed	Gravel

Piping

Main piping size	4"
Main piping material	Saran lined
Main valving arrangement	Nest of Auto. valves

TABLE 53A PRESSURE FILTERS (continued)

Control System

Control	Automatic
Initiation of regeneration	Time Clock
Backwash control	Limit stop on valves

Auxiliaries

Interconnecting piping between multiple units	4" wrought steel & saran lined
Pressure gauges	3 pr included
Sample cocks	3 pr

Electrical Requirements

Volts, Hertz, Phase	115/60/1

ADDITIONAL SPECIFICATIONS:

Bypass type rate of flow meters, 2 on each unit	Included
Waste discharge interconnecting headers	4" saran lined
Filters are backwashed with raw AMD water. Valve nest, waste lines, and AMD supplies are corrosion resistant for the service	Included

TABLE 54

MISCELLANEOUS ITEMS
DETAILED SPECIFICATIONS, 0.5 MGD

Air compressor for all plant control needs	Included
Modulating pH meter, alarm, pH flow cell and electrodes with milliamps output to control pH correction pump. Also strip chart recorder	Included
Acid feed pump with variable speed drive modulated by milliamp signal from pH meter	Included
Horizontal acid storage tank, black steel construction, unlined with breather for 66°Be H_2SO_4. Designed for concrete saddles which are not included	5' dia. x 8' long
Acid supply and feed piping from storage tank to pump and point of feed	½" carpentier 20SS
All interplant piping needed to interconnect ion exchange units, aerator, clarifier and filters in saran lined and black wrought steel as required	Included

* *

The following items are specifically excluded.

1. The building, its foundations, concrete reservoir, concrete clarifier tank, and concrete tank saddles.
2. Auxiliary plumbing and plumbing fixtures for the building.
3. Electrical wiring of building and connections between electrical controls.
4. All pump starters.
5. Installation and erection of equipment.
6. The equipment to handle builk unslaked lime is not considered to be included as part of this plant design.

NOTE: The handling and supply of the large quantities of carbon dioxide required would well be the subject of a separate study. The reader should realize that other methods of supply might be usuable and should conduct his own evaluation.

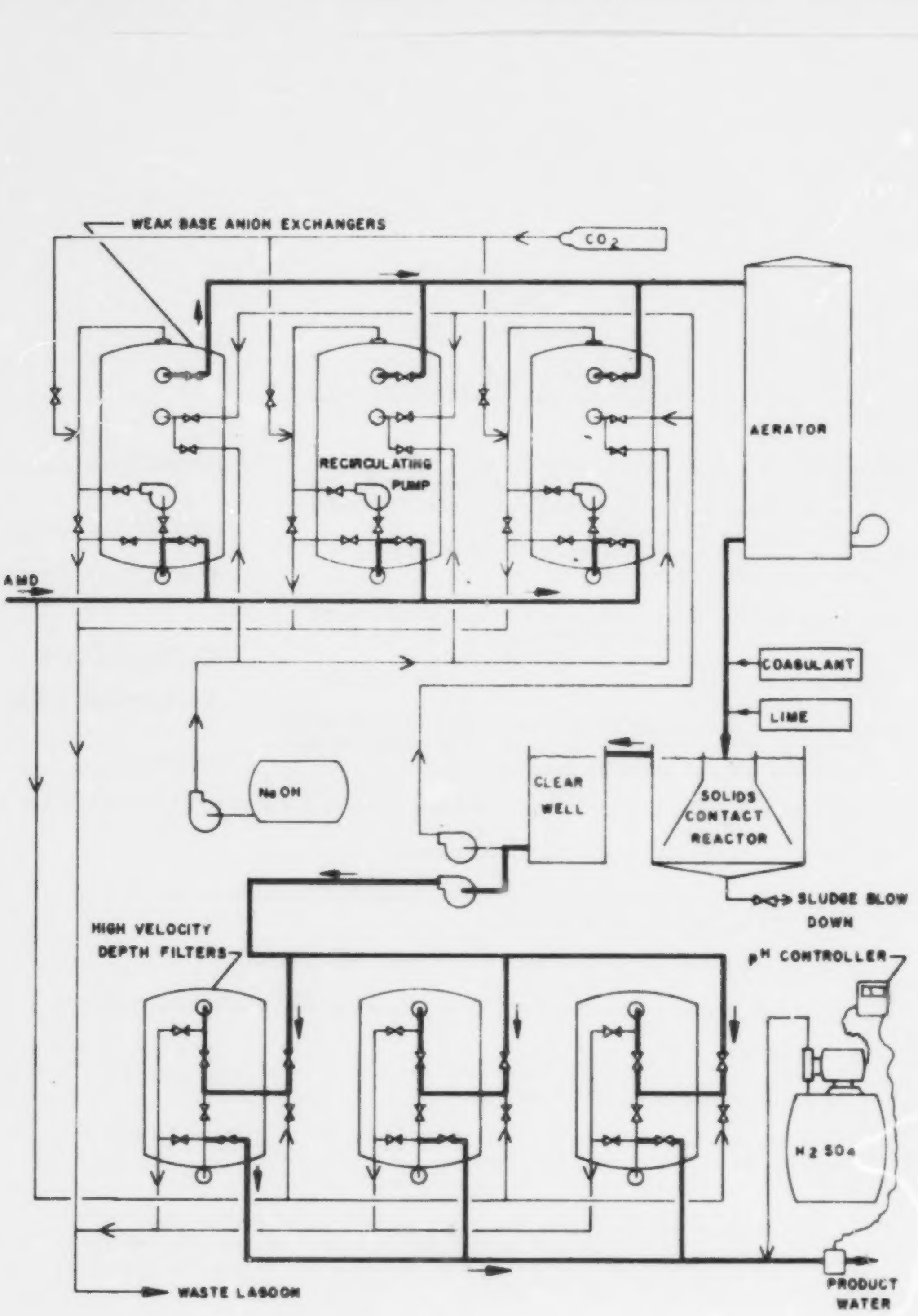
WEAK BASE ANION EXCHANGERS
CO_2
AERATOR
RECIRCULATING PUMP
AMD
COAGULANT
LIME
NaOH
CLEAR WELL
SOLIDS CONTACT REACTOR
SLUDGE BLOW DOWN
HIGH VELOCITY DEPTH FILTERS
pH CONTROLLER
H_2SO_4
WASTE LAGOON
PRODUCT WATER

FIGURE 27. AMD Treatment Plant Flow Diagram, Modified Desal System 0.5 MGD

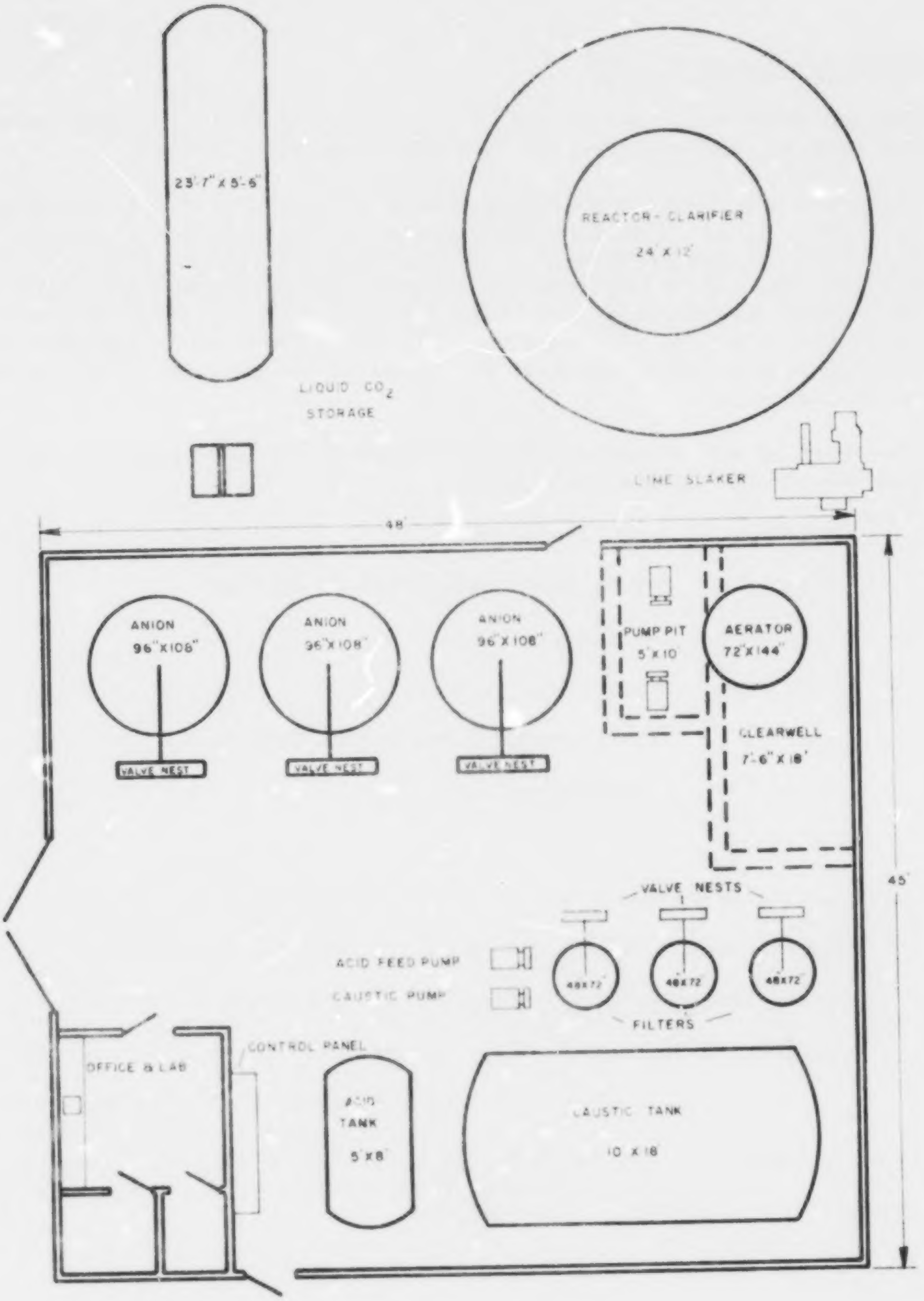
23'-7" X 5'-6"
REACTOR - CLARIFIER
24' X 12'
LIQUID CO_2
STORAGE
LIME SLAKER
48'
45'
ANION
96" X 108"
ANION
96" X 108"
ANION
96" X 108"
VALVE NEST
VALVE NEST
VALVE NEST
PUMP PIT
5' X 10'
AERATOR
72" X 144"
CLEARWELL
7'-6" X 18'
VALVE NESTS
ACID FEED PUMP
CAUSTIC PUMP
48X72
48X72
48X72
FILTERS
OFFICE & LAB
CONTROL PANEL
ACID
TANK
5' X 8'
CAUSTIC TANK
10' X 18'

FIGURE 28. AMD Treatment Plant Plan, Modified Desal System 0.5 MGD

Plant to produce 1.0 MGD:

The equipment will consist essentially of three anion exchange vessels, one aerator, one reactor-clarifier and three final filters.

Three anion exchange vessels are incorporated in the design although only one will be in service operation at a time. Three are required because total regeneration time of one vessel is longer than its service run. The complete cycle for one tank is three hours service, four hours regeneration, two hours standby. Thus, if only one tank is on stream at a time and takes the full plant flow, eight regenerations per day is a workable schedule and produces the required plant output of treated water.

The detailed specifications of the equipment for this plant to treat 1.0 MGD are on the following pages.

Table 55	Anion Exchangers
Table 56	Forced Draft Degasifier or Aerator
Table 57	Reactor-Clarifier
Table 58	Pressure Filters
Table 59	Miscellaneous Items and Exclusions
Figure 29	Flow Diagram
Figure 30	Plant Plan

Chemical operating costs are estimated as follows:

Caustic Soda, 50%, 11,904 lb/day, $0.037 per pound = $440.45

Carbon Dioxide, 9,821 lb/day, $0.015 per pound = 147.32

Lime, 3,500 lb/day, $0.01 per pound = 35.00

Total, $/day = $622.87

Cost estimates for the equipment as specified in Tables 55 through 59 for this 1.0 MGD size plant, excluding freight, building and land, and assembly and erection, total $465,000.

Cost estimates for erection have been made as follows:

Electrical	$ 9,300
Plumbing	50,000

TABLE 55
ANION EXCHANGERS

DETAILED SPECIFICATIONS, 1.0 MGD

PERFORMANCE:

Total System

Total influent exchangeable anions, gpg as $CaCO_3$	84.0
Design flow rate, gpm	695
Operating water pressure, psig	40, minimum
Number of units	Three

Per Unit

Design flow rate, gpm	738
Peak flow rate, gpm	738
Backwash rate, gpm	155
Anion exchange material, type	IRA-68
quantity, cu ft	372
capacity, Kgr per cu ft	30
capacity, Kgr per unit	11,160
Gallons treated per regeneration	132,855
Gallons net to service	125,100
Regenerant quantity per regeneration, lbs.	1,488
Sodium Hydroxide	1,228
Carbon Dioxide	

SPECIFICATIONS:

Model Number	None

Tanks

Tank diameter	120"
Straight side of tank	126"
Design working pressure of tank	100 psi Non-Code
External surface	Prime Painted
Tank lining, material and thickness	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" x 16" manhole

Internals

Backwash collecting manifold, design & materials	PVC header lateral
Treated water collecting manifold, design & materials	PVC-Screened header lateral
Regenerant distributor, design and materials	PVC header lateral
Underdrain system, design and materials	PVC header lateral
Supporting bed	Silica gravel

Table continued on next page

TABLE 55A ANION EXCHANGERS (continued)

Piping

Main piping size	6" x 4"
Main piping material	Saran lined steel
Main valving arrangement	Nest of auto diaph type
Main valving material	Saran lined cast iron

Control System

Control	Automatic
Initiation of regeneration	pH meter
Backwash control	Limit stop on valves

Auxiliaries

Meter, size and type	None
Meter register	--
Interconnecting piping between multiple units, inlet and outlet	8" saran lined
Pressure gauges	chemical seal type
Sample cocks	1 pair per unit
Conductivity instrument, type	None
manufacturer	--
model number	--

Regeneration Equipment

Type of regenerant introduction	Positive displace. pump
Regenerant introduction strength	4%
Regenerant tank size, horizontal bulk storage	10' dia x 34'
Material of construction	Unlined steel

Electrical Requirements

Volts, Hertz, Phase	115/60/1 & 230-460/60/3

ADDITIONAL SPECIFICATIONS:

Regenerant caustic pump, Milton Roy, 2HP	610 gph
Regenerant caustic piping system	1" wrought steel
Regeneration water header from clear well	4" wrought steel
Regeneration water pump-Gould #3196 all iron centrifugal for 175 gpm @ 50' head with 5 HP ODP motor	Included
Waste discharge piping	4" saran lined steel
Bypass type rate of flow meters, 2 for each tank	4"
Bypass type rate of flow meter, 1 for product output	8"
Beckman Model 940 pH meter with flow chamber, and electrodes to detect endpoints	1 for each tank
Three pen strip chart recorder	4"
25 ton liquid CO_2 storage container with 2 HP compressor, 18 KW vaporizer and regulator	Included

Table continued on next page

TABLE 55B ANION EXCHANGERS (continued)

3" stainless steel sparger to dispurse CO_2 gas in recirculating line	1 per tank
Pressure relief and suction relief on each tank	Included
Piping from CO_2 storage to points of use, wrought steel	Included
Pump to recycle water during CO_2 saturation, Gould #3196 all stainless steel, 25 HP, 3600 rpm, 230-460/60/3 ODP motor for 186 gpm @ 100 psi	1 for each tank

TABLE 56
FORCED DRAFT DEGASIFIERS OR AERATOR
DETAILED SPECIFICATIONS, 1.0 MGD

PERFORMANCE:

Total System

Function	Aeration
Influent iron and manganese content, ppm maximum	140
Influent temperature, °F	Ambient
Design flow rate, gpm	695
Water pressure at inlet	15 psi
Number of units	One

Per Unit

Design flow rate, gpm	738
Peak flow rate, gpm	738

SPECIFICATIONS:

Tower

Size of tower	96" diameter
Height of tower	144"
Materials of construction	Fir Staves

Internals

Material of packing	Red wood tray
Depth of packing	9 feet
Inlet distributor, design and materials	PVC header lateral
Support, design and materials	Wood

Auxiliaries

Blower, type three units needed	Centrifugal
capacity, cfm	3200 each
static head, inches H_2O	2
Motor, type	2 HP - ODP
voltage, current, phases	230-460/60/3
Level control, type for clear well	Modulating "levetrol"
Inlet valve, type	Modulated
material of contruction	Cast iron - SS trim
size	4"

STORAGE: None. Water flows directly to Reactor-Clarifier and from there to concrete clear well. Aerator is located above clear well. Water level in clear well controls aerator inlet valve and thus also feed to reactor-clarifier. Modulating level control is itemized above.

TABLE 57

REACTOR-CLARIFIER
DETAILED SPECIFICATION, 1.0 MGD

Size of Tank, diameter x side wall depth	32' x 13'
Tank construction	Concrete
Design flow rate, gpm	738
Rise rate, gpm per sq. ft.	1
Detention period, minutes	102
Bottom slope to central sludge cone	1 in 12
Provisions for mixing, internal recirculation, flocculation, settling, clarification, positive sludge thickening and removal	Included
Lime slaker and coagulant feeder	Included
Sludge rake drive mechanism fully enclosed with motor	1.0 HP
Turbine driving mechanism with motor	3.0 HP
Beam type superstructure spanning tank	36" wide floor plate
Conical reaction chamber	Included
Peripheral effluent collection launder	Included
Sampling pipes	Included
Plug type sludge valve with pneumatic cylinder operator	Included
Piping from aerator to reactor-clarifier, welded	8" - Included
Piping from reactor-clarifier to clear well, welded	8" - Included
Estimated lime dosage, lbs. of 100% CaO per 1000 gal.	3.2
lbs. of 100% CaO per day	3,401

Clarifier will be mostly below ground, only the top 18" to 24" of the concrete tank will be avoe ground level.

CLEAR WATER

Concrete construction, below floor level of equipment building, size 9.5' x 25' x 10' deep. Nominal capacity 14,000 gallons. Concrete construction is not included as part of equipment design cost and should be included under building construction. Two pumps are located in the pump pit adjacent to clear well. One is for regenerating the anion units and is described under Anion Exchangers. The other is a transfer or forwarding pump for regular service operation. It is a Gould #3196 all iron centrifugal pump for 700 gpm at 50 psi with 40 HP, 230-460/60/3 1750 rpm ODP motor.

TABLE 58

PRESSURE FILTERS
DETAILED SPECIFICATIONS, 1.0 MGD

PERFORMANCE:

Total System

Design flow rate, gpm	695 gpm
Operating water pressure, psig	30 psi min.
Number of units	Three
Type of units	Hi-Velocity

Per Unit

Design flow rate, gpm	233
Peak flow rate, gpm	350
Design rate, gpm/ft.^{2} of filter area	10
Backwash rate, gpm	350
Filter media, type	Hi-Velocity
quantity, cu. ft.	65
depth of bed, inches	33

SPECIFICATIONS:

Model Number	None

Tanks

Tank diameter	66"
Straight side of tank	84"
Design working pressure of tank	100 psi Non-Code
External surface	Prime painted
Internal surface	3/32" Plastisol
Tank supports	Adjustable jacks
Access opening(s)	12" x 16" manhole

Internals

Inlet distributor, design and materials	PVC Header lateral
Underdrain system, design and materials	PVC Header lateral
Supporting bed	Gravel

Piping

Main piping size	4" x 6"
Main piping material	Saran lined
Main valving arrangement	Nest of Auto. valves

Table continued on next page

TABLE 58A PRESSURE FILTERS (continued)

Control System	
Control	Automatic
Initiation of regeneration	Time Clock
Backwash control	Limit stop on valves
Auxiliaries	
Interconnecting piping between multiple units	6" pr steel
Pressure gauges	3 pr incluced
Sample cocks	3 pr
Electrical Requirements	
Volts, Hertz, Phase	115/60/1

ADDITIONAL SPECIFICATIONS:

Bypass type rate of flow meters, 2 on each unit	Included
Waste discharge interconnecting headers	6" saran lined
Filters are backwashed with raw AMD water. Valve nest, waste lines, and AMD supplies are corrosion resistant for the service	Included

TABLE 59

MISCELLANEOUS ITEMS
DETAILED SPECIFICATIONS, 1.0 MGD

Air compressor for all plant control needs	Included
Modulating pH meter, alarm, pH flow cell and electrodes with milliamps output to control pH correction pump. Also strip chart recorder	Included
Acid feed pump with variable speed drive modulated by milliamp signal from pH meter	Included
Horizontal acid storage tank, black steel construction, unlined with breather for 66°Be H_2SO_4. Designed for concrete saddles which are not included.	6' dia x 12' long
Acid supply and feed piping from storage tank to pump and point of feed.	½" carpentier 20SS
All interplant piping needed to interconnect ion exchange units, aerator, clarifier and filters in saran lined and black wrought steel as required	Included

* * * * * * * * * * * * * * * * * * * *

The following items are specifically excluded.

1. The building, its foundations, concrete reservoir, concrete clarifier tank, and concrete tank saddles.
2. Auxiliary plumbing and plumbing fixtures for the building.
3. Electrical wiring of building and connections between electrical controls.
4. All pump starters.
5. Installation and erection of equipment.
6. The equipment to handle bulk unslaked lime is not considered to be included as part of this plant design.

NOTE: The handling and supply of the large quantities of carbon dioxide required would well be the subject of a separate study. The reader should realize that other methods of supply might be usable and should conduct his own evaluation.

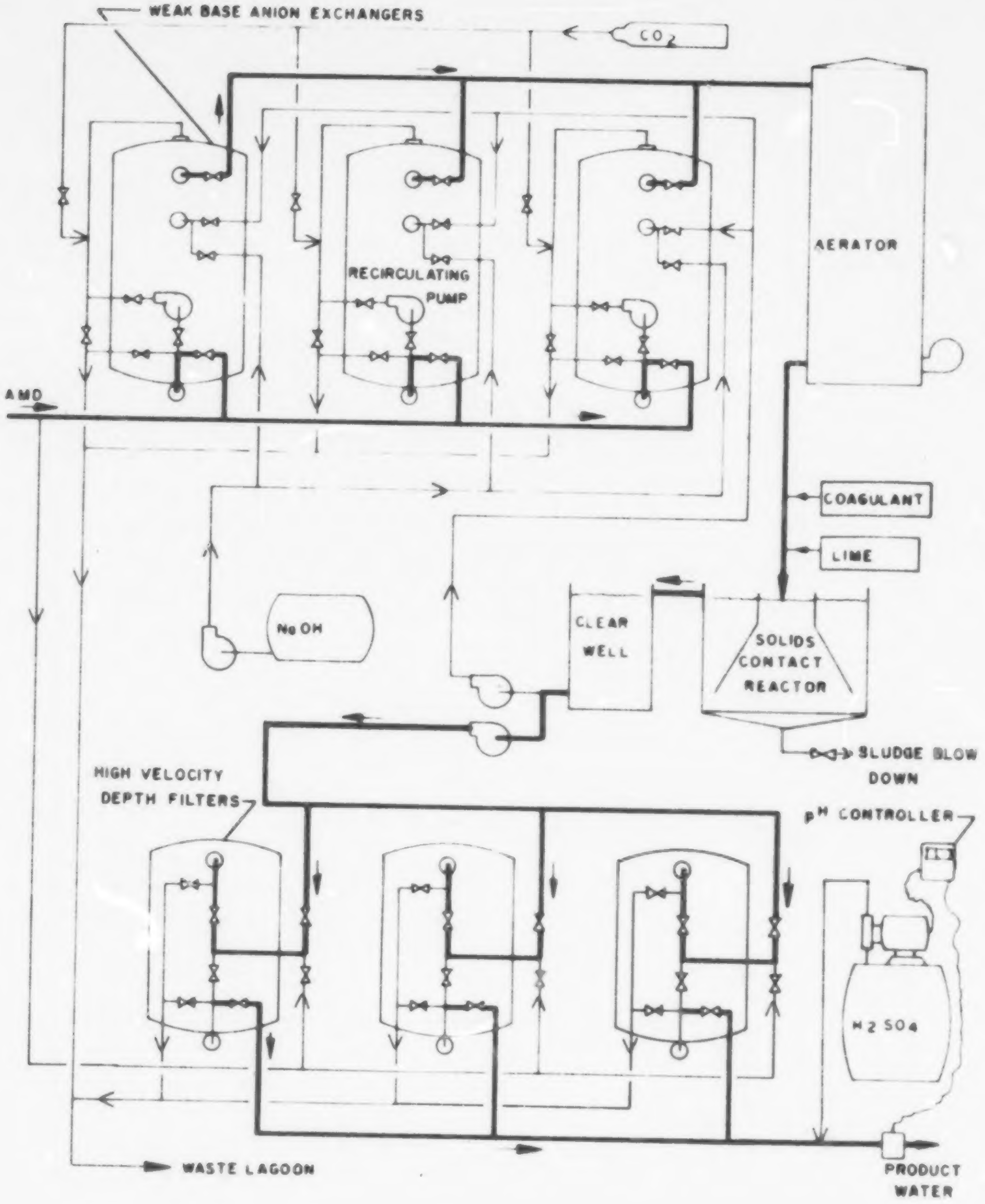
WEAK BASE ANION EXCHANGERS
CO2
AERATOR
RECIRCULATING PUMP
AMD
COAGULANT
LIME
NaOH
CLEAR WELL
SOLIDS CONTACT REACTOR
SLUDGE BLOW DOWN
HIGH VELOCITY DEPTH FILTERS
pH CONTROLLER
H2SO4
WASTE LAGOON
PRODUCT WATER

FIGURE 29. AMD Treatment Plant Flow Desal System 1.0 MGD

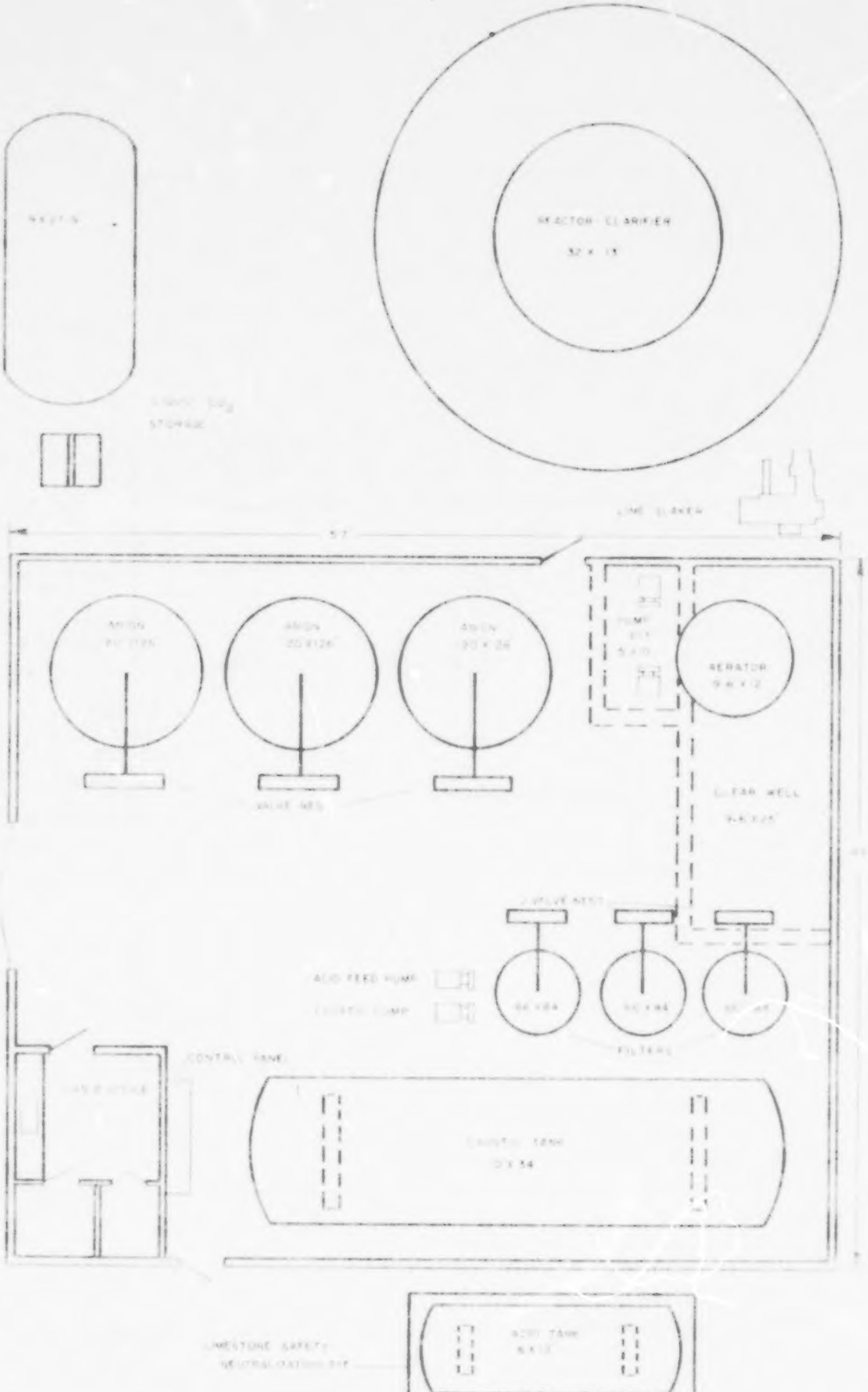
REACTOR-CLARIFIER
32 × 15
LIQUID SO2
STORAGE
LIME SLAKER
57'
ANION
20 × 26
ANION
20 × 26
ANION
20 × 26
PUMP
PIT
5 × 10
AERATOR
9.6 × 12
VALVE NES
CLEAR WELL
9.6 × 25
VALVE NEST
ACID FEED PUMP
CAUSTIC PUMP
96 × 84
96 × 84
96 × 84
FILTERS
CONTROL PANEL
LAB & OFFICE
CAUSTIC TANK
12 × 34
LIMESTONE SAFETY
NEUTRALIZATION PIT
ACID TANK
6 × 12

FIGURE [illegible] AMD Treatment Plant Plan, Modified Desal System 1.0 MGD

SECTION 18

ACKNOWLEDGEMENTS

The authors, Jim Holmes and Ed Kreusch, gratefully acknowledge the varied assistance received from many sources in the completion of this project. Financial support was received from the Water Quality Office, Environmental Protection Agency. The guidance of the Project Officer, Mr. Ron Hill of the National Environmental Research Center, Cincinnati, Ohio was a firm foundation for the project's inception. Dr. James Shackelford from the Washington office of the Environmental Protection Agency was helpful in consultation about the progress as it was achieved.

Cost estimates for the erection of the plants were obtained from separate organizations. Their help in furnishing these erection estimates is thankfully recognized. The cost estimates for the electrical installations were furnished by the following company:

Hyre Electric Company
Mr. Frank Farrel, Chief Field Engineer
Chicago, Illinois 60608

The cost estimates for the plumbing installation were furnished by the following company:

Associated Piping Contractors, Inc.
Carmen Perna, President
Chicago, Illinois 60656

The services of Farouk Husseini and Willard Rakow were valuable in performing the laboratory tests with the ion exchange column. Messrs. Dick Bezjian and James O'Malley were helpful in providing the analyses of the numerous varied samples collected during the laboratory studies. Mr. Don Senger performed a significant portion of the work for this project by furnishing the designs and specifications for the six treatment plants. Mr. Larry Coshenet provided the drafting services which were required. Other services of separate departments of the parent company, Culligan International Company, are gratefully recognized. Their assistance, which was beyond their responsibilities in support of the commercial organization, embraced areas which were beyond the fields of specialization of the authors.

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SECTION 19

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SECTION 20

DEFINITIONS

ACIDITY	Acidity is the capacity to donate hydrogen ions. The acidity is normally expressed (in the ion exchange industry) in terms of calcium carbonate equivalents. Acidity is quantitatively measured by titration to selected endpoints with a standard solution-usually of sodium hydroxide. The donation of hydrogen ions to depress the pH below 8.3 is termed total acidity: below pH 4.3 is "free mineral acidity". (see also FMA.)
ALKALINITY	Capacity to neutralize acids. In water, most alkalinity is due to the water's content of bicarbonates, carbonates, or hydroxide. The alkalinity is normally expressed in terms of calcium carbonate equivalents.
ANION	An ionic particle which is negatively charged.
BACKWASH	Reverse (normally upwards) flow through a bed of mineral or ion exchange resin to remove insoluble particulates and to loosen the bed.
BED DEPTH	The height of mineral, or ion exchange resin in a column.
BED EXPANSION	The amount of expansion given to a bed of mineral or ion exchange resin, by upflow passage of water. It is usually expressed as a percent of the unexpanded bed.
BED VOLUME	The amount of mineral, or ion exchange resin, in a column.
BREAKTHROUGH	Refers to the concentration of a particular ion, or other substance in the effluent from a treatment system. Breakthrough occurs when the effluent concentration rapidly increases. Normally, when the breakthrough concentration reaches about 10% of the influent concentration, exhaustion has occurred.

CALCIUM CARBONATE EQUIVALENT	An expression for the concentration of constituents on a common basis for ease of calculation. Conversion of the quantity expressed "as calcium carbonate" to "as another form" requires multiplication by the ratio of the chemical equivalent weight of the desired form to that of calcium carbonate. For example, 80 mg/l of magnesium as calcium carbonate becomes 44.4 mg/l (80 x 12.2/20) as magnesium.
CAPACITY	The quantitative ability of a treatment component or system to perform. With ion exchange systems, this quantity is expressed as kilograins per cubic foot.
CATION	An ionic particle which is positively charged.
COCURRENT	Operation of a column of ion exchange resin or other mineral, with the service cycle and the regeneration cycle performed in the same direction, both either upflow or downflow.
COMPOSITE SAMPLE	A sample collected to be representative of a water flow which continues for an extended period of time.
CONDUCTIVITY	Ability of water to conduct electricity; it is the reciprocal of resistivity. Conductivity is measured in reciprocal ohms per centimeter. Water with a low concentration of ionic solids will have very low conductivity.
COUNTERCURRENT	Operation of a column of ion exchange resin or other mineral, with the service cycle and the regeneration cycle performed in opposite directions.
DEMINERALIZATION	Reduction of the ionic content of water.
DOWNFLOW	Direction of flow of solutions through ion exchange, or mineral bed columns during operation; in at the top and out at the bottom of the column.

DRIP SAMPLE	A composite sample collected by slow continuous sampling of a flowing stream.
EFFLUENT	The solution which emerges from a component or system.
ELUATE	Effluent during regeneration of an ion exchange resin. (See "Elution").
ELUENT	Influent regeneration solution to an ion exchange resin. (See "Elution").
ELUTION	The removal of an adsorbed ion or ions from an ion exchange resin during regeneration by using solutions containing relatively high concentrations of other ions. This latter solution is called the eluant. During elution, the eluant removes the adsorbed ions from the ion exchange resin; the effluent solution which contains the eluted ions is then called the eluate.
ENDPOINT	The achievement of exhaustion. With ion exchange resins, the endpoint of the service cycle is at 10% breakthrough.
EXHAUSTION CYCLE	The function of a process component in the service cycle. The regenerated form of a weak base resin without adsorbed acids.
FMA FREE MINERAL ACIDITY	Strong acids, which in water are formed principly by chloride or sulfate ions when the water has been treated by a cation exchange resin in the hydrogen form.
gpg GRAINS PER GALLON	A unit of concentration (weight per volume) that is used in the ion exchange industry. (See "GRAIN".) One gpg is numerically equal to 17.1 mg/l.
GRAIN	A unit of weight, being numerically equal to 1/7000th of a pound. (See "GRAINS PER GALLON".)

gpm	Gallons per minute.
gpm/cu ft	Gallons per minute per cubic foot of ion exchange resin or other mineral in a column.
gpm/sq ft	Gallons per minute per square foot of cross-sectional area.
HARDNESS	The sum of the calcium and magnesium ions. although other polyvalent cations are included at times. Hardness is normally expressed in terms of calcium carbonate equivalents.
ION EXCHANGE RESIN	An insoluble material which can remove ions by replacing them with an equivalent amount of a similarly charged ion.
kgr KILOGRAINS	A unit of weight (1.000 grains) equal to 1/7th of a pound.
kgr/cu ft	Kilograins (expressed as calcium carbonate) per cubic foot of ion exchange resin.
LEAKAGE	The amount of unadsorbed ion present in the effluent of a treatment component.
LIME	Lime refers to compounds of calcium. Hydrated lime is calcium hydroxide. Lime which is not hydrated is referred to as quick lime, which is calcium oxide.
MICROMHOS	Unit of measurement of electrical conductivity.
mg/l MILLIGRAMS PER LITER	A unit of concentration referring to the milligrams weight of a solute per liter of solution. The term is approximately equal to the older "part per million" term.
NEUTRALIZATION	Mutual reaction of acids and alkalies until the concentrations of hydrogen and hydroxyl ions in solution are at the desired value which is usually approximately equal.

ppm PARTS PER MILLION	A unit of concentration, which in the water treatment industry equals one part of solute in one million parts by weight of solvent. It is approximately equal to the more precise term mg/l.
REGENERANT	A solution of relatively high ionic concentration used to restore an ion exchange resin to its desired ionic form.
REGENERATION	Restoration of an ion exchange resin to its desired ionic form.
RINSE	The removal of excess regenerant from an ion exchange resin.
SALT SPLITTING	The conversion of neutral salts to their corresponding acids or bases.
SERVICE CYCLE	The use of a process component to perform its desired function.
SLUDGE	Settled precipitates of large amount.
SLUDGE BLANKET	A layer of sludge which is suspended by upflow passage of water.
SOFTENING	Removal of the hardness (calcium and magnesium ions) from water.
UPFLOW	Direction of flow of water upwardly through a component.
WEAK ACID RESIN	A cation exchange resin which cannot split neutral salts.
WEAK BASE RESIN	An anion exchange resin which cannot split neutral salts, but will merely absorb free mineral acidity.

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SECTION 21

APPENDIX

COLUMN EFFLUENT ANALYSES

TABLE 60
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 2A (Cation H^+)

Endpoint

Date: 8/19/70

Run No.										
Throughput, gals/cu ft	0	153	268							
Ferrous, ppm Fe	22.4	28.0	258							
Ferric, ppm Fe	1.6	1.0	29							
Calcium, ppm Ca	15.0	15.0	164							
Magnesium, ppm Mg	2.9	3.7	34							
Aluminum, ppm Al										
Manganese, ppm Mn	0.9	1.1	9.8							
Sodium, ppm Na	4.0	52	51							
FMA, ppm $CaCO_3$	1430	925	415							
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1470	1100	1600							
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.80	2.05	2.45							
Sp Resistance, ohm-cm	130	190	305							
Temperature, °F	67	67	67							
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 61

ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE

Column Effluent Analyses

Run 4A (Cation H⁺)

Endpoint

Date: 8/21/70

Run No.										
Throughput, gals/cu ft	0	82	170	247						
Ferrous, ppm Fe	2.8	2.8	2.8	42						
Ferric, ppm Fe	0.9	0.7	0.9	1.0						
Calcium, ppm Ca	7.0	5.6	5.5	28.1						
Magnesium, ppm Mg	0.6	0.5	0.6	6.8						
Aluminum, ppm Al										
Manganese, ppm Mn	0.2	0.2	0.2	1.8						
Sodium, ppm Na	0.6	0.6	8.3	39						
FMA, ppm $CaCO_3$	1620	1668	1652	1452						
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1500	1420	1370	1540						
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.75	1.75	1.70	1.80						
Sp Resistance, ohm-cm	120	120	120	120						
Temperature, °F	68.5									
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 62
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 8A (Cation H$^+$)

Endpoint

Date: 9/1/70

Run No.										
Throughput, gals/cu ft	0	60	120	182	242					
Ferrous, ppm Fe	2.8	2.8	2.8	2.8	46					
Ferric, ppm Fe	1.5	2.3	2.0	2.0	2.0					
Calcium, ppm Ca	7.3	5.8	5.5	5.5	29					
Magnesium, ppm Mg	0.9	0.8	0.8	0.8	8.0					
Aluminum, ppm Al										
Manganese, ppm Mn	0	0	0	0	2.0					
Sodium, ppm Na	0.4	0.4	0.4	0.7	1.8					
FMA, ppm $CaCO_3$	1500		1560	1564	1430					
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1470		1520	1530	1560					
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.70		1.70	1.70	1.75					
Sp Resistance, ohm-cm	120		120	115	130					
Temperature, °F	64.0	65.0	65.5	65.5	65.0					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 63
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 12A (Cation H^+)

Endpoint Date: 9/8/70

Run No.										
Throughput, gals/cu ft	0	183	212	242	274	304				
Ferrous, ppm Fe	1.0	0	0	1.0	3.0	28				
Ferric, ppm Fe	1.1	1.9	1.9	1.2	1.8	0				
Calcium, ppm Ca	3.8	4.9	2.5	2.6	3.6	15				
Magnesium, ppm Mg	0.3	0.3	0.2	0.3	0.6	4.1				
Aluminum, ppm Al	0.5	0	0.5	0	0.5	0.5				
Manganese, ppm Mn	0.1	0.1	0.1	0.1	0.2	1.2				
Sodium, ppm Na	0	0	0	0	0.1	0.2				
FMA, ppm $CaCO_3$	1580	1580			1560	1540				
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1530	1530			1510	1590				
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.73	1.73			1.73	1.75				
Sp Resistance, ohm-cm	115	110			115	115				
Temperature, °F	77.0	77.0	77.0	77.0	77.0	77.0				
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 64
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 19A (Cation H^+)

Endpoint Date: 9/18/70

Run No.										
Throughput, gals/cu ft.	0	212	242	272	304					
Ferrous, ppm Fe	1.0	0	1.0	3.0	22					
Ferric, ppm Fe	1.0	2.2	1.4	1.6	3					
Calcium, ppm Ca	4.0	3.0	3.0	4.0	14					
Magnesium, ppm Mg	0.3	0.3	0.3	0.7	4.5					
Aluminum, ppm Al	0	0	0	0	0.5					
Manganese, ppm Mn	0.1	0.1	0.1	0.2	1.2					
Sodium, ppm Na	0	0	0.2	0.2	0.3					
FMA, ppm $CaCO_3$	1500				1560					
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1450				1420					
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.70				1.70					
Sp Resistance, ohm-cm	120				115					
Temperature, °F	74.0	74.0	74.0	74.0	74.0					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 65
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 4B (Cation H^+)

Endpoint

Date: 8/21/70

Run No.										
Throughput, gals/cu ft	0	77	153	230						
Ferrous, ppm Fe	0	0	8.4	165						
Ferric, ppm Fe	1.0	0.9	1.0	32						
Calcium, ppm Ca	2.6	2.3	7.1	103						
Magnesium, ppm Mg	0.2	0.3	1.6	26						
Aluminum, ppm Al										
Manganese, ppm Mn	0.1	0.1	0.5	6.6						
Sodium, ppm Na	0.4	1.7	15	35						
FMA, ppm $CaCO_3$	1120	1500	1400	1400						
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1090	1450	1420	2060						
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.85	1.70	1.70	1.70						
Sp Resistance, ohm-cm	160	115	115	120						
Temperature, °F	68.5									
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 66
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 8B (Cation H^+)

Endpoint

Date: 9/1/70

Run No.										
Throughput, gals/cu ft	0	60	120	180	210					
Ferrous, ppm Fe	0	0	0	8.4	115					
Ferric, ppm Fe	1.0	0.5	1.1	2.0	11					
Calcium, ppm Ca	2.1	1.1	1.1	5.4	62					
Magnesium, ppm Mg	0.4	0.2	0.2	1.7	18					
Aluminum, ppm Al										
Manganese, ppm Mn	0	0	0	0	6.0					
Sodium, ppm Na	0.4	0.4	0.4	0.8	1.6					
FMA, ppm $CaCO_3$	1456		1540	1540	1235					
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1410		1480	1520	1610					
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.73		1.70	1.70	1.85					
Sp Resistance, ohm-cm	125		115	115	150					
Temperature °F	64.0	65.0	65.5	65.5	65.0					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 67
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 12B (Cation H^+)

Endpoint

Date: 9/8/70

Run No.										
Throughput, gals/cu ft	0	180	210	240						
Ferrous, ppm Fe	0	0	5.0	14						
Ferric, ppm Fe	0.8	1.5	0	0						
Calcium, ppm Ca	1.8	1.5	2.4	6.7						
Magnesium, ppm Mg	0.1	0.2	0.7	2.0						
Aluminum, ppm Al	0.5	0	0	0						
Manganese, ppm Mn	0.1	0.1	0.3	0.6						
Sodium, ppm Na	0	0.1	0.1	0.1						
FMA, ppm $CaCO_3$	1600	1600		1600						
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1540	1540		1580						
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.70	1.73		1.74						
Sp Resistance, ohm-cm	110	110		110						
Temperature, °F	77	77		77						
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 68
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 19B (Cation H+)

Endpoint

Date: 9/18/70

Run No.										
Throughput, gals/cu ft	0	212	242	272	302					
Ferrous, ppm Fe	0	0	0	8.0	28					
Ferric, ppm Fe	0.6	1.3	3.0	0.4	1.0					
Calcium, ppm Ca	2.0	1.0	2.0	6.0	13					
Magnesium, ppm Mg	0.1	0.2	0.5	1.5	5.2					
Aluminum, ppm Al	0	0	0	0	0					
Manganese, ppm Mn	0	0	0.1	0.4	1.3					
Sodium, ppm Na	0.1	0.1	0.1	0.1	0.2					
FMA, ppm $CaCO_3$	1620				1580					
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1560				1620					
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.65				1.70					
Sp Resistance, ohm-cm	110				115					
Temperature, °F	74	74	74	74	74					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 69
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 23A (Cation Na$^+$)

Endpoint

Date: 10/1/70

Run No.										
Throughput, gals/cu ft	120	240	360	420	450	480				
Ferrous, ppm Fe	0	0	0	2.0	4.0	18				
Ferric, ppm Fe	3.3	2.7	1.8	1.1	2.0	1.0				
Calcium, ppm Ca	6.7	4.6	3.3	4.2	5.2	11				
Magnesium, ppm Mg	0.4	0.4	0.2	0.5	1.1	3.3				
Aluminum, ppm Al	0.1	0.3	0.2	0.4	0.3	0.3				
Manganese, ppm Mn	0.1	0.1	0.1	0.1	0.2	0.7				
Sodium, ppm Na	735	525	360	320	330	290				
FMA, ppm $CaCO_3$	0	465	815	890	895	920				
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1560	1560	1450	1540	1570	1570				
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	4.20	2.30	2.05	2.00	2.00	1.98				
Sp Resistance, ohm-cm	320	225	180	170	170	165				
Temperature, °F	74	74	74	74	74	74				
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 70
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 26A (Cation Na⁺)

Endpoint

Date: 10/9/70

Run No.										
Throughput, gals/cu ft	120	240	300	360	420	480				
Ferrous, ppm Fe	0	0	0	0	5.0	25				
Ferric, ppm Fe	2.3	2.3	2.3	2.9	3.4	0				
Calcium, ppm Ca	8.6	5.7	5.0	5.0	8.8	23				
Magnesium, ppm Mg	0.3	0.2	0.2	0.3	1.4	5.2				
Aluminum, ppm Al	0.1	0.0	0.0	0.0	0.0	0.1				
Manganese, ppm Mn	0.1	0.1	0.1	0.1	0.3	1.2				
Sodium, ppm Na	720	500	410	360	310	280				
FMA, ppm $CaCO_3$	15	495	675	815	865	900				
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1540	1540	1520	1550	1500	1600				
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	3.75	2.30	2.15	2.00	1.97	1.97				
Sp Resistance, ohm-cm	330	225	195	180	175	175				
Temperature, °F	73	73	73	73	73	73				
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 71
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 30A (Cation Na+)

Endpoint Date: 10/16/70

Run No.										
Throughput, gals/cu ft	120	240	360	420	450					
Ferrous, ppm Fe	0	0	0	8.0	25					
Ferric, ppm Fe	1.4	1.8	2.1	1.0	2.0					
Calcium, ppm Ca	2.7	2.4	2.0	4.3	9.6					
Magnesium, ppm Mg	0.4	0.3	0.3	1.9	6.1					
Aluminum, ppm Al	0	0.1	0.1	0.1	0.2					
Manganese, ppm Mn	0.2	0.1	0.2	0.5	1.5					
Sodium, ppm Na	660	500	360	320	310					
FMA, ppm $CaCO_3$	10		795	855	875					
Hot FMA, ppm $CaCO_3$				915	940					
Sulfate, ppm SO_4	1217		927	1537	1397					
Chloride, ppm Cl	130		130	130	130					
Alkalinity, ppm $CaCO_3$										
pH	3.94		2.10	2.05	2.06					
Sp Resistance, ohm-cm	320		170	165	165					
Temperature, °F	72	72	72	72	72					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 72
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 23B (Cation Na$^+$)

Endpoint

Date: 10/1/70

Run No.										
Throughput, gals/cu ft	120	215	240	330	360	390				
Ferrous, ppm Fe	8.0	6.0	4.0	4.0	14	43				
Ferric, ppm Fe	1.0	1.2	1.8	1.9	0	1.0				
Calcium, ppm Ca	19	15	13	15	15	38				
Magnesium, ppm Mg	1.4	1.1	0.9	1.0	2.2	8.0				
Aluminum, ppm Al	0.1	0.1	0	0.1	0.1	0.1				
Manganese, ppm Mn	0.4	0.3	0.2	0.2	0.5	1.5				
Sodium, ppm Na	655	515	400	300	280	245				
FMA, ppm $CaCO_3$	65	420	700	900	930	920				
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	1500	1520	1550	1540	1550	1600				
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	3.24	2.36	2.14	2.00	1.98	1.98				
Sp Resistance, ohm-cm	310	230	190	165	165	165				
Temperature, °F	74	74	74	74	74	74				
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 73
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 39A (Wk. Base NH_2)

Date: 11/17/70

Run No.										
Throughput, gals/cu ft	58	178	530	815	970					
Ferrous, ppm Fe	0	3.0	87	101	104					
Ferric, ppm Fe	0.7	4.1	5.0	7.0	3.0					
Calcium, ppm Ca	199	226	205	207	204					
Magnesium, ppm Mg	15	38	31	31	31					
Aluminum, ppm Al	0.1	0	7.8	22	28					
Manganese, ppm Mn	0.0	3.1	7.8	7.9	7.9					
Sodium, ppm Na	14	7.0	4.9	5.9	5.6					
FMA, ppm $CaCO_3$	0	0	0	0	5					
Hot FMA, ppm $CaCO_3$					465					
Sulfate, ppm SO_4	385	460	500	540	540					
Chloride, ppm Cl	131	179	207	202	202					
Alkalinity, ppm $CaCO_3$	8	8	6	4	0					
pH	5.30	5.30	4.65	4.40	4.05					
Sp Resistance, ohm-cm	960	800	720	670	650					
Temperature, °F	72	73	73	73	73					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 74
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 34B (Wk. Base NH_2)

Date: 11/2/70

Run No.										
Throughput, gals/cu ft	57	227	455	550	680	790	850			
Ferrous, ppm Fe	3.0	62	73	87	92	98	104			
Ferric, ppm Fe	20	40	15	10	20	15	14			
Calcium, ppm Ca	233	209	199	209	206	201	208			
Magnesium, ppm Mg	37	30	30	30	30	30	30			
Aluminum, ppm Al	1.3	2.8	4.2	5.4	11	17	19			
Manganese, ppm Mn	0.3	8.9	8.2	8.2	8.1	8.2	8.2			
Sodium, ppm Na	10	1.5	1.2	6.4	1.4	1.2	1.4			
FMA, ppm $CaCO_3$	0	0	0	0	0	0	5			
Hot FMA, ppm $CaCO_3$							395			
Sulfate, ppm SO_4	505	400	432	515	505	516	545			
Chloride, ppm Cl	136	179	196	199	202	202	202			
Alkalinity, ppm $CaCO_3$	44	110	42	16	0	0	0			
pH	7.70	6.30	5.55	4.85	4.50	4.20	4.00			
Sp Resistance, ohm-cm	800	780	770	750	720	700	670			
Temperature, °F	73	73	73	73	73	73	73			
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 75
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 39B (Wk. Base NH_2)

Date: 11/17/70

Run No.										
Throughput, gals/cu ft	124	378	496	600	720					
Ferrous, ppm Fe	0	81	81	87	90					
Ferric, ppm Fe	1.0	12	11	9	7					
Calcium, ppm Ca	238	213	207	202	203					
Magnesium, ppm Mg	35	33	31	31	31					
Aluminum, ppm Al	0.0	2.4	11	16	18					
Manganese, ppm Mn	0.8	7.8	7.8	7.8	7.8					
Sodium, ppm Na	11	13	6.5	7.0	6.6					
FMA, ppm $CaCO_3$	0	0	0	0	5					
Hot FMA, ppm $CaCO_3$					385					
Sulfate, ppm SO_4	495	560	525	515	530					
Chloride, ppm Cl	173	193	193	199	199					
Alkalinity, ppm $CaCO_3$	6	4	2	2	0					
pH	5.70	4.90	4.55	4.30	4.00					
Sp Resistance, ohm-cm	790	740	730	710	700					
Temperature, °F	72	73	73	73	73					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 76
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 49A (Strong Base, $SO_4^=$) - 21K with limed AMD Regenerant

Date: 1/6/71

	Endpoint					Influent AMD				
Run No.										
Throughput, gals/cu ft	0	21	27	33						
Ferrous, ppm Fe	64	98	101	101		104				
Ferric, ppm Fe	2	32	19	19		106				
Calcium, ppm Ca	155	188	185	185		184				
Magnesium, ppm Mg	24	28	29	29		29				
Aluminum, ppm Al	0.3	14	19	20		15				
Manganese, ppm Mn	5.8	7.3	7.4	7.6		7.7				
Sodium, ppm Na	1.2	1.1	0.8	0.8		0.7				
FMA, ppm $CaCO_3$	0	0	10	45		535				
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	410	545	535	535		1010				
Chloride, ppm Cl	119	148	156	156		196				
Alkalinity, ppm $CaCO_3$	20	0	0	0		0				
pH	4.95	4.30	3.70	3.20		2.40				
Sp Resistance, ohm-cm	880	720	670	610		320				
Temperature, °F	72	72	72	72						
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 77
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 50A (Strong Base $SO_4^=$) - IRA-410 with limed AMD Regenerant

Endpoint Influent AMD Date: 1/12/71

Run No.										
Throughput, gals/cu ft	0	21								
Ferrous, ppm Fe	95	101		98						
Ferric, ppm Fe	15	9		102						
Calcium, ppm Ca	180	180		170						
Magnesium, ppm Mg	28	28		28						
Aluminum, ppm Al	0.0	17		15						
Manganese, ppm Mn	7.3	7.8		7.5						
Sodium, ppm Na	0.5	0.4		0.9						
FMA, ppm $CaCO_3$	0	20		540						
Hot FMA, ppm $CaCO_3$				840						
Sulfate, ppm SO_4	515	490		955						
Chloride, ppm Cl	145	159		196						
Alkalinity, ppm $CaCO_3$	10	0		0						
pH	4.80	3.40		2.35						
Sp Resistance, ohm-cm	780	640		350						
Temperature, °F	72	72								
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 78

ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE

Column Effluent Analyses

Run 52A (Strong Base $SO_4^=$) - IRA-410 with NH_4OH regenerant

Date: 1/14/71

Run No.					Endpoint				Influent	AMD
Throughput, gals/cu ft	0	21	41	62	70	79				
Ferrous, ppm Fe	25	42	50	95	98	101			104	
Ferric, ppm Fe	21	1.0	3.0	0	2.0	5.0			96	
Calcium, ppm Ca	200	200	197	192	192	180			201	
Magnesium, ppm Mg	26	26	28	27	29	29			28	
Aluminum, ppm Al	0.0	0.2	0.7	6.7	26	31			15	
Manganese, ppm Mn	7.0	7.9	7.9	7.9	7.9	7.8			7.9	
Sodium, ppm Na	0.8	1.7	0.6	0.5	0.9	0.4			1.7	
FMA, ppm $CaCO_3$	0	0	0	0	0	30			545	
Hot FMA, ppm $CaCO_3$				255		480				
Sulfate, ppm SO_4	360	350	370	570	575	570			1048	
Chloride, ppm Cl	88	91	94	128	136	142			196	
Alkalinity, ppm $CaCO_3$	170	205	195	15	0	0			0	
pH	5.90	6.05	6.10	4.80	4.25	3.50			2.40	
Sp Resistance, ohm-cm	940	900	870	770	700	640			330	
Temperature, ° F										
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 79
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 55A (Strong Base $SO_4^=$) - IRA-410 with NH_4OH regenerant

Date: 1/20/71

Run No.						Endpoint				Influent AMD
Throughput, gals/cu ft	0	21	31	41	52	58	68			
Ferrous, ppm Fe	0	0	20	78	98	101	101			104
Ferric, ppm Fe	0.6	0.5	0	8.0	4.0	4.0	15			114
Calcium, ppm Ca	21	210	207	198	209	195	196			210
Magnesium, ppm Mg	0	0	33	31	28	28	28			27
Aluminum, ppm Al	0	0	0.8	1.6	7.6	18	27			18
Manganese, ppm Mn	0	0	25	12	9.0	9.0	9.0			8.4
Sodium, ppm Na	0.2	0.5	0.5	0.8	0.3	0.4	0.4			0.3
FMA, ppm $CaCO_3$	0	0	0	0	0	0	25			860
Hot FMA, ppm $CaCO_3$						345	430			1205
Sulfate, ppm SO_4	33	370	530	580	640	610	630			1360
Chloride, ppm Cl	9	94	97	102	111	116	119			201
Alkalinity, ppm $CaCO_3$	5	10	45	40	5	2.5	0			0
pH	8.20	8.70	6.20	6.00	4.60	4.35	3.65			2.15
Sp Resistance, ohm-cm	3900	880	830	810	760	740	690			215
Temperature, °F	70	70	70	70	70	70	70			
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 80

ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE

Column Effluent Analyses

Run 49B (Strong Base $SO_4^{=}$) 21K with NH_4OH Regenerant

Date: 1/6/71

Run No.				Endpoint			Influent	AMD		
Throughput, gals/cu ft	0	21	43	49	55					
Ferrous, ppm Fe	0	59	98	101	101		104			
Ferric, ppm Fe	0.4	2.0	22	19	29		106			
Calcium, ppm Ca	92	182	180	183	191		184			
Magnesium, ppm Mg	17	26	28	28	29		29			
Aluminum, ppm Al	0	0	15	28	29		15			
Manganese, ppm Mn	0	7.0	7.9	8.3	8.6		7.7			
Sodium, ppm Na	1.6	3.7	4.0	0.7	5.6		0.7			
FMA, ppm $CaCO_3$	0	0	0	0	20		535			
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	203	530	635	635	680		1010			
Chloride, ppm Cl	48	68	74	80	82		196			
Alkalinity, ppm $CaCO_3$	25	40	--	--	--		0			
pH	6.10	5.20	4.25	4.20	3.70		2.40			
Sp Resistance, ohm-cm	1200	850	710	700	690		320			
Temperature, °F	72	72	72	72	72					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 81
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 55B (Strong Base $SO_4^{=}$) 21K with NH_4OH regenerant

Endpoint

Date: 1/20/71

Run No.								Influent	AMD	
Throughput, gals/cu ft	0	17	21	31	35	43				
Ferrous, ppm Fe	0	74	81	101	104	104		104		
Ferric, ppm Fe	0.2	3	8	2	11	17		114		
Calcium, ppm Ca	20	228	201	195	200	200		210		
Magnesium, ppm Mg	0	39	30	30	31	32		27		
Aluminum, ppm Al	0.3	0.3	0.5	16	26	25		18		
Manganese, ppm Mn	0	12	8.6	8.6	8.8	9.0		8.4		
Sodium, ppm Na	0.3	0.5	1.3	0.5	0.4	1.3		0.3		
FMA, ppm $CaCO_3$	0	0	0	0	10	60		860		
Hot FMA, ppm $CaCO_3$					400	455		1205		
Sulfate, ppm SO_4	40	695	615	640	680	730		1360		
Chloride, ppm Cl	3	82	82	94	94	97		201		
Alkalinity, ppm $CaCO_3$	5	45	35	5	0	0		0		
pH	7.90	5.80	5.50	4.50	3.95	3.30		2.15		
Sp Resistance, ohm-cm	2000	800	790	730	690	620		215		
Temperature, °F	70	70	70	70	70	70				
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 82
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 64A (Cation H+)

Date: 2/11/71

Run No.										
Throughput, gals/cu ft	41	227	360	370	380					
Ferrous, ppm Fe	3.7	1.3	3.7	5.7	9.0					
Ferric, ppm Fe	0.7	0	0	0.8	1.6					
Calcium, ppm Ca	4.5	2.5	3.2	4.0	5.4					
Magnesium, ppm Mg	0.6	0.1	0.4	0.9	1.5					
Aluminum, ppm Al	1	0	0	0	0					
Manganese, ppm Mn	0.1	0.0	0.1	0.2	0.3					
Sodium, ppm Na	0.2	0.1	1.5	1.4	1.7					
FMA, ppm $CaCO_3$	1500				1480					
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4										
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.70				1.70					
Sp Resistance, ohm-cm	115				120					
Temperature, °F	68									
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 83
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 69A (Cation H+)

Date: 2/19/71

Run No.										
Throughput, gals/cu ft	41	227	302	310						
Ferrous, ppm Fe	3.9	2.1	9.0	14.8						
Ferric, ppm Fe	1.2	0.3	3.5	7.2						
Calcium, ppm Ca	4.8	3.1	8.0	12.7						
Magnesium, ppm Mg	0.6	0.2	2.0	3.7						
Aluminum, ppm Al	0	0	0	0						
Manganese, ppm Mn	0.1	0.0	0.5	0.9						
Sodium, ppm Na	1.0	0.2	0.9	1.1						
FMA, ppm $CaCO_3$	1660			1600						
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4										
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.75			1.75						
Sp Resistance, ohm-cm	115			120						
Temperature, °F	70									
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 84
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 74A (Cation H^+)

Date: 2/26/71

Run No.										
Throughput, gals/cu ft	41	123	227	302	309					
Ferrous, ppm Fe	4.8	3.3	3.1	9.6	13					
Ferric, ppm Fe	0.1	0.1	0	0.4	3					
Calcium, ppm Ca	5.4	4.0	3.9	7.5	8.8					
Magnesium, ppm Mg	0.6	0.3	0.3	1.5	2.2					
Aluminum, ppm Al	0	0	0	0	0					
Manganese, ppm Mn	0.1	0.1	0.1	0.3	0.5					
Sodium, ppm Na	3.2	1.5	3.8	38	45					
FMA, ppm $CaCO_3$	1500				1360					
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4										
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.75				1.80					
Sp Resistance, ohm-cm	120				130					
Temperature, °F	71									
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 85
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 77B (Cation H^+)

Date: 3/3/71

Run No.										
Throughput, gals/cu ft	41	123	223	230						
Ferrous, ppm Fe	25	14	22	31						
Ferric, ppm Fe	1	3	2	2						
Calcium, ppm Ca	25	17	18	22						
Magnesium, ppm Mg	3.3	2.2	3.2	4.6						
Aluminum, ppm Al	0	0	0	0						
Manganese, ppm Mn	0.9	0.5	0.8	1.1						
Sodium, ppm Na	0.3	0.0	2.1	2.3						
FMA, ppm $CaCO_3$	1400			1380						
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4										
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.80			1.80						
Sp Resistance, ohm-cm	125			130						
Temperature, °F	70									
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 86
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 86A (Cation H+)

Date: 3/24/71

Run No.										
Throughput, gals/cu ft	38	115	203	210	230					
Ferrous, ppm Fe	0.5	0.9	0.6	0.8	2.0					
Ferric, ppm Fe	8.5	4.0	12.4	18.2	31					
Calcium, ppm Ca	35	23	41	52	80					
Magnesium, ppm Mg	4.5	2.8	8.1	12	18					
Aluminum, ppm Al	0	0	0	0	0					
Manganese, ppm Mn	1.2	0.7	2.0	2.6	4.4					
Sodium, ppm Na	0.3	0.2	1.2	1.2	1.6					
FMA, ppm $CaCO_3$	1540			1440	1320					
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4										
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.70			1.75	1.80					
Sp Resistance, ohm-cm	100			109	120					
Temperature, °F	70									
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 87
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Run 86B (Cation H+)

Date: 3/24/71

Run No.										
Throughput, gals/cu ft	38	118	176	185						
Ferrous, ppm Fe	0.3	0.3	0.5	1.2						
Ferric, ppm Fe	1.4	1.3	8.1	18.8						
Calcium, ppm Ca	18	13	27	48						
Magnesium, ppm Mg	2.2	1.8	6.5	16						
Aluminum, ppm Al	0	0	0	0						
Manganese, ppm Mn	0.5	0.5	1.6	3.4						
Sodium, ppm Na	0.1	0.6	1.3	1.3						
FMA, ppm $CaCO_3$	1580			1400						
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4										
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.70			1.75						
Sp Resistance, ohm-cm	99			110						
Temperature, °F	70									
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 88
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Runs 94 thru 98 (Composite Cation, Weak Base, Lime, Filter Effluent) 100% Ferrous

Date: 4/13/71 thru 4/29/71

Run No.	94	95	96	97	98					
Throughput, gals/cu ft	260	400	420	420	420					
Ferrous, ppm Fe	0.0	0.0	0.0	0.0	0.0					
Ferric, ppm Fe	0.1	0.0	0.1	0.0	0.0					
Calcium, ppm Ca	21	18	13	20	17					
Magnesium, ppm Mg	0.9	1.5	1.7	1.6	1.5					
Aluminum, ppm Al	0	0	0	0	0					
Manganese, ppm Mn	0.00	0.17	0.17	0.05	0.04					
Sodium, ppm Na	0.6	8.9	44	23	23					
FMA, ppm $CaCO_3$	0	0	0	0	0					
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	31	43	110	64	61					
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$	25	25	22.5	40	35					
pH	10.00	9.50	7.60	9.95	9.90					
Sp Resistance, ohm-cm	5900	6400	3400	4400	4900					
Temperature, °F										
Loading Factor, gpg $CaCO_3$	83.0	84.0	83.0	85.0	83.0					
Capacity, Kgrs/cu ft	21.6	33.6	34.8	35.6	34.8					

TABLE 89

ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE

Column Effluent Analyses

Runs 99 thru 107 (Composite Cation, Weak Base, Lime, Filter Effluent) 100% Ferric

Date: 5/5/71 thru 6/1/71

Run No.	99	100	101	102	103	104	105	106	107	
Throughput, gals/cu ft	370	360	340	360	360	360	360	340	320	
Ferrous, ppm Fe	0.1	0.1	0.0	0.1	0.0	0.0	0.0	0.0	0.0	
Ferric, ppm Fe	0.0	0.0	0.2	0.1	0.0	0.0	0.0	0.0	0.0	
Calcium, ppm Ca	21	23	17	21	22	18	18	26	11	
Magnesium, ppm Mg	2.1	1.8	2.1	2.1	2.2	2.2	2.1	2.3	0.2	
Aluminum, ppm Al	0	0	0	0	0	0	0	0	0	
Manganese, ppm Mn	0.00	0.01	0.01	0.01	0.02	0.01	0.00	0.00	0.00	
Sodium, ppm Na	27	26	14	24	35	55	39	61	200	
FMA, ppm $CaCO_3$	0	0	0	0	0	0	0	0	0	
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	86	83	60	80	101	129	95	160	274	
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$	30	35	20	30	35	40	40	40	180	
pH	10.05	9.85	9.90	10.15	9.90	10.00	10.10	10.00	11.40	
Sp Resistance, ohm-cm	3800	4100	5300	4100	3200	3100	3200	2500	810	
Temperature, °F										
Loading Factor, gpg $CaCO_3$	102.0	103.0	105.0	102.0	107.0	103.0	105.0	106.0	105.0	
Capacity, Kgrs/cu ft	37.7	37.1	35.7	36.7	38.5	37.2	37.8	36.0	33.6	

TABLE 90
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Runs 108 thru 112 (Composite Cation, Weak Base, Lime, Filter Effluent) 100% Ferric

Date: 6/2/71 thru 6/14/71

Run No.	108	109	110	111	112					
Throughput, gals/cu ft	320	320	320	366	340					
Ferrous, ppm Fe	0.1	0.0	0.1	0.0	0.1					
Ferric, ppm Fe	0.0	0.2	0.0	0.0	0.4					
Calcium, ppm Ca	9.5	23	15	14	22					
Magnesium, ppm Mg	2.1	3.9	2.4	2.3	2.5					
Aluminum, ppm Al	0	0	0	0	0					
Manganese, ppm Mn	0.01	0.02	0.01	0.01	0.00					
Sodium, ppm Na	100	57	43	65	43					
FMA, ppm $CaCO_3$	0	0	0	0	0					
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	184	165	103	136	153					
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$	60	25	35	45	20					
pH	10.30	9.90	10.00	10.00	10.00					
Sp Resistance, ohm-cm	1800	2400	3600	2600	3100					
Temperature, °F										
Loading Factor, gpg $CaCO_3$	106.0	106.0	108.0	110.0	112.0					
Capacity, Kgrs/cu ft	33.9	33.9	34.5	40.2	38.0					

TABLE 91
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Runs 94 thru 98 (Composite Strong Acid Cation Effluent) 100% Ferrous

Date: 4/13/71 thru 4/29/71

Run No.	94	95	96	97	98					
Throughput, gals/cu ft										
Ferrous, ppm Fe	5.5	7.4	11.4	11.4	10.2					
Ferric, ppm Fe	1.1	2.6	2.6	1.6	1.8					
Calcium, ppm Ca	13	12	15	15	13					
Magnesium, ppm Mg	1.5	1.8	2.0	1.7	1.5					
Aluminum, ppm Al	0	0	0	0	0					
Manganese, ppm Mn	0.48	0.59	0.59	0.54	0.47					
Sodium, ppm Na	0.3	5.7	18	14	8.7					
FMA, ppm $CaCO_3$	1440	1440	1420	1460	1420					
Hot FMA, ppm $CaCO_3$	1580	1595	1545	1545	1540					
Sulfate, ppm SO_4	1430	1430	1430	1470	1420					
Chloride, ppm Cl	--	--	--	--	--					
Alkalinity, ppm $CaCO_3$	--	--	--	--	--					
pH	1.70	2.20	1.80	1.80	1.85					
Sp Resistance, ohm-cm	120	120	120	120	120					
Temperature, °F	69	70	70	70	70					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 92

ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE

Column Effluent Analyses

Runs 99 thru 107 (Composite Strong Acid Cation Effluent) 100% Ferric

Date: 5/5/71 thru 6/1/71

Run No.	99	100	101	102	103	104	105	106	107	
Throughput, gals/cu ft										
Ferrous, ppm Fe	13	5.1	1.7	1.2	0.4	0.4	0.3	0.2	0.2	
Ferric, ppm Fe	2.0	2.2	2.4	2.4	3.1	3.1	2.9	2.7	3.2	
Calcium, ppm Ca	21	21	19	21	20	21	20	19	18	
Magnesium, ppm Mg	2.3	2.3	2.3	2.4	2.5	2.5	2.4	2.4	2.5	
Aluminum, ppm Al	0	0	0	0	0	0	0	0	0	
Manganese, ppm Mn	0.76	0.76	0.79	0.85	0.87	0.84	0.77	0.75	0.80	
Sodium, ppm Na	8.2	4.5	8.3	8.9	8.9	5.6	2.9	5.3	4.6	
FMA, ppm $CaCO_3$	1760	1780	1800	1760	1840	1780	1800	1820	1800	
Hot FMA, ppm $CaCO_3$	2000	1980	2000	1980	2000	2020	2020	2020	2060	
Sulfate, ppm SO_4	1770	1780	1930	1900	1830	1770	1790	1800	1780	
Chloride, ppm Cl	--	--	--	--	--	--	--	--	--	
Alkalinity, ppm $CaCO_3$	--	--	--	--	--	--	--	--	--	
pH	1.80	1.75	1.75	1.65	1.75	1.65	1.65	1.70	1.75	
Sp Resistance, ohm-cm	100	98	98	98	97	98	98	98	98	
Temperature, °F	70	70	70	70	70	70	70	70	70	
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 93
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Runs 108 thru 112 (Composite Strong Acid Cation Effluent) 100% Ferric

Date: 6/2/71 thru 6/14/71

Run No.	108	109	110	111	112					
Throughput, gals/cu ft										
Ferrous, ppm Fe	0.2	0.2	0.2	0.2	0.2					
Ferric, ppm Fe	3.1	4.8	3.5	2.9	3.8					
Calcium, ppm Ca	14	27	16	20	21					
Magnesium, ppm Mg	2.6	4.1	2.6	2.6	2.8					
Aluminum, ppm Al	0	0	0	0	1.0					
Manganese, ppm Mn	0.85	1.3	0.81	0.86	0.90					
Sodium, ppm Na	15	5.3	6.9	21	13					
FMA, ppm $CaCO_3$	1820	1820	1860	1880	1920					
Hot FMA, ppm $CaCO_3$	1980	1960	1980	2040	2020					
Sulfate, ppm SO_4	1800	1830	1840	1860	1910					
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$										
pH	1.80	1.70	1.70	1.85	1.80					
Sp Resistance, ohm-cm	98	96	98	98	96					
Temperature, °F	70	70	70	70	70					
Loading Factor, gpg $CaCO_3$										
Capacity, Kgrs/cu ft										

TABLE 94
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Runs 113 thru 117 and 122 (Weak Base HCO_3^- - Lime) 100% Ferrous

Date: 6/17/71 thru 7/1/71

Run No.	113	114	115	116	117	122				
Throughput, gals/cu ft	595	440	465	444	448	450				
Ferrous, ppm Fe	0.0	0	0	0	0	0				
Ferric, ppm Fe	0.0	0	0	0	0	0				
Calcium, ppm Ca	62	30	10	49	21	17				
Magnesium, ppm Mg	3.0	2.2	1.9	0.3	1.2	1.5				
Aluminum, ppm Al	0	0	0	0	0	0				
Manganese, ppm Mn	0.01	0.01	0.00	0.02	0.00	0.01				
Sodium, ppm Na	0.9	0.9	0.7	1.4	1.1	1.3				
FMA, ppm $CaCO_3$	0	0	0	0	0	0				
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	133	53	5	63	28	15				
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$	25	30	30	60	30	35				
pH	10.00	10.40	10.10	10.40	10.00	10.1				
Sp Resistance, ohm-cm	2700	4400	5000	2500	4500	4700				
Temperature, °F										
Loading Factor, gpg $CaCO_3$	64.0	70.5	69.5	70.0	67.5	68.5				
Capacity, Kgrs/cu ft	33.0	31.0	32.4	31.1	30.3	30.8				

TABLE 95
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Runs 123 thru 128 (Weak Base HCO_3^- - Lime) 100% Ferric

Date: 7/21/71 thru 8/2/71

Run No.	123	124	125	126	127	128				
Throughput, gals/cu ft	260	281	260	281	194	226				
Ferrous, ppm Fe	0.1	0.0	0.0	0.0	0.0	0.0				
Ferric, ppm Fe	0.0	0.0	0.0	0.0	0.0	0.0				
Calcium, ppm Ca	18	39	7.9	110	110	16				
Magnesium, ppm Mg	2.8	0.2	1.2	0	0.1	9.0				
Aluminum, ppm Al	1.0	1.0	0	2.0	1.0	0				
Manganese, ppm Mn	0.01	0.00	0.01	0.01	0.00	0.00				
Sodium, ppm Na	1.7	2.1	2.1	2.0	2.4	2.0				
FMA, ppm $CaCO_3$	0	0	0	0	0	0				
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	30	27	0	13	58	54				
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$	30	75	30	265	220	25				
pH	10.30	10.20	10.00	10.80	11.60	9.90				
Sp Resistance, ohm-cm	6500	2600	6500	850	1000	6400				
Temperature, °F										
Loading Factor, gpg $CaCO_3$	96.5	96.0	97.0	95.0	94.0	96.5				
Capacity, Kgrs/cu ft	25.1	27.0	25.2	26.7	18.2	21.8				

TABLE 96
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Runs 118 thru 121 (Weak Base HCO_3^- - Weak Acid H+) 100% Ferrous

Date: 7/6/71 thru 7/19/71

Run No.	118	119	120	121						
Throughput, gals/cu ft	460	404	356	440						
Ferrous, ppm Fe	0.0	0.0	0.0	0.0						
Ferric, ppm Fe	0.2	0.0	0.2	0.0						
Calcium, ppm Ca	19	12	21	17						
Magnesium, ppm Mg	1.4	3.3	0.6	0.3						
Aluminum, ppm Al	0	0	0	1						
Manganese, ppm Mn	0.03	0.01	0.00	0.03						
Sodium, ppm Na	2.4	1.4	1.4	1.1						
FMA, ppm $CaCO_3$	0	0	0	0						
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	22	11	11	6						
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$	35	25	45	40						
pH	9.70	9.80	10.30	10.90						
Sp Resistance, ohm-cm	9300	8400	6000	5600						
Temperature, °F										
Loading Factor, gpg $CaCO_3$	68.5	69.5	64.0	57.5						
Capacity, Kgrs/cu ft	31.5	28.1	22.8	25.2						

TABLE 97
ION EXCHANGE TREATMENT OF ACID MINE DRAINAGE
Column Effluent Analyses

Runs 129 thru 135 (Weak Base HCO_3^- - Weak Acid - Lime) 100% Ferric

Date: 8/4/71 thru 8/20/71

Run No.	129	130	131	132	133	134	135			
Throughput, gals/cu ft	240	250	248	242	260	260	260			
Ferrous, ppm Fe	0	0	0	0	0	0	0			
Ferric, ppm Fe	0	0	0	0	0	0	0			
Calcium, ppm Ca	10	11	9.1	5.5	5.6	11	6.5			
Magnesium, ppm Mg	1.0	0.2	0.1	0.3	0.4	0.3	0.5			
Aluminum, ppm Al	0	0	0	0	0	0.2	0			
Manganese, ppm Mn	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Sodium, ppm Na	0.1	0.1	0.2	0.4	0.5	0.5	0.4			
FMA, ppm $CaCO_3$	0	0	0	0	0	0	0			
Hot FMA, ppm $CaCO_3$										
Sulfate, ppm SO_4	4	9	0	0	0	4	0			
Chloride, ppm Cl										
Alkalinity, ppm $CaCO_3$	25	20	23	16	16	25	18			
pH	9.80	10.00	10.00	10.00	9.90	9.80	9.90			
Sp Resistance, ohm-cm	12,500	16,000	17,000	15,000	16,000	15,000	13,000			
Temperature, °F										
Loading Factor, gpg $CaCO_3$	95.0	90.0	90.5	86.5	88.0	38.5	87.0			
Capacity, Kgrs/cu ft	22.8	22.5	22.5	21.0	21.6	23.0	22.6			

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